BUILDING A BRIDGE TO THE CORN ETHANOL INDUSTRY

CORN STOVER TO ETHANOL AT HIGH PLAINS CORPORATION'S YORK, NEBRASKA CO-LOCATED PLANT SITE

FINAL REPORT
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1 EXECUTIVE SUMMARY

The United States Department of Energy (DOE) Office of Fuel Development OFD supports the commercialization of lignocellulose (fibrous plant matter) to fuel ethanol. The majority of the work that has received this support is focused on the development of the technologies, which will make this goal a reality. The technologies available as a result of this work are making fuel ethanol from lignocellulose a more feasible option for our energy future. However, the capital required to obtain the economies of scale at a greenfield site are cost prohibitive at this time. In an effort to minimize the cost, and operational difficulties associated with a greenfield site, DOE – through the National Renewable Energy Laboratory (NREL) – has turned to the corn-to-ethanol industry.

The corn-to-ethanol industry is responsible for nearly the entire U.S. supply of fuel ethanol. With its years of experience in industrial scale fuel ethanol operations and infrastructure, the corn-to-ethanol industry could play an important role in the initiation of a lignocellulose-to-ethanol industry. In addition to its experience and infrastructure, the locations of these plants are generally located in the heart of corn country, where there is an ample supply of low cost lignocellulose…corn stover.

The purpose of this project and report was to investigate the co-location of a corn stover-to-ethanol facility at High Plains Corporation's York, Nebraska facility. It was hoped that this co-location strategy would allow the stover facility to operate with less overhead cost, less operations costs, and lower capital cost through infrastructure sharing with the existing corn facility. This is as opposed to a greenfield located stover-to-ethanol facility. Although the result of this configuration did not turn out to be economicly attractive, we identified issues which, when the solutions are found, could produce positive economics.

The lignocellulosic technology chosen was based on the NREL lignocellulosic biomass to ethanol process[10] using dilute sulfuric acid pretreatment. This is followed by separate enzymatic hydrolysis, using cellulase enzyme, and co-fermentation by the recombinant bacterium *Zymonomas mobilis* developed at NREL. The enzyme is produced on site using an enzyme production technology from Pure Vision Technology, Inc., which results in a higher specific activity (more effective) enzyme than the lignocellulosic reference model. The *Z. mobilis* is capable of fermenting both five and six carbon sugars.

The scale of the facility was determined by data gathered by High Plains Corp. as to the amount of stover available in a reasonable harvest area around the existing York, NE plant. This 900 dry metric ton per day (347,223 short ton/yr) of stover resulted in a scale-down of the NREL reference model to 45% based on feedstock throughput. The resulting stover plant produces 25,746,124 gallons per year of fuel grade ethanol, which is 97.7% of the theoretical 45% scale down. This is not 100% of the theoretical scale down due to a slightly lower conversion efficiency of cellulose to glucose that results from separate hydrolysis and fermentation occurring in a much shorter period of time than in the reference model.

Although this facility has a less efficient hydrolysis, the hydrolysis and fermentation are accomplished in 57% of the time. This trade-off was accepted with the intention of reducing capital and operating costs that result from shorter residence times via smaller or fewer vessels. The resulting yield is 81.7 gallons of ethanol per dry metric ton of corn stover (74 gal. per short ton). This is a slightly lower yield than the 83.5 gallons per dry metric ton (76 gal. per short ton) reported for the lignocellulosic model. The benefits of reduced hydrolysis and fermentation time result from the use of the higher specific activity enzyme and the separation of hydrolysis and fermentation (as opposed to Simultaneous Saccharification and Co-Fermentation – SSCF).

Appendix 5 (the equipment list) has comparisons between the study equipment costs and the reference model as scaled to 45% (1999 costs with area weighted average scaling exponents used). Also included is a comparison of the electrical workloads. The workloads and equipment costs are organized by area. The comparison shows that the co-located study model equipment costs are $14.8 million less than the lignocellulosic model. This represents a 19.5% cost savings. There appears to be a $0.5 million capital savings by separating hydrolysis and co-fermentation. This is a result of fewer vessels required due to decreased residence time as noted in Table 6.2.5.A. The use of the PureVision cellulase production technology appears to result in a $1.7 million capital cost savings due to the reduced cellusase prduction scale required as a result of the higher specific activity of the enzyme. Installation factors have been revised (in most cases increased) for the co-located study case and this has an effect on the installed costs (see Volume II of this report). A comparison of the installation factors and weighted average scaling exponents is also on the equipment list under each area.

Under the assumptions of this project, there is no collection and market of CO_2. Although most of the CO_2 currently produced at the High Plains Corp. York facility is marketed to a CO_2 compressing company located on adjacent property, there appears to be no further interest at this time in marketing of additional CO_2 in this way.

The detoxification of stover slurry before hydrolysis and fermentation produces a significant amount of gypsum waste (over 60,000 lbs/day). This will incur a disposal cost and the facility would benefit greatly from either the elimination of its production or the development of a market for the low quality gypsum.

Eight railcars are filled with high water content lignin waste each day. The marketing of this waste as an energy-containing co-product is critical to the economics of the facility. If 10% of the water were separated from the lignin waste and sent to waste water treatment, it is likely that there would be some wastewater discharge to the city of York to reduce salts (under current design there is no waste water discharge to the city with the exception of treated storm run-off water). This would increase the wet fuel value of the lignin on a per ton basis.

As a result of this study, several critical issues were brought to light. The most important is the development of a system for feedstock harvest, transport, storage and processing. The very large volume of low-density biomass will require bulk-handling methods. Bales (the existing supply model) are too cumbersome for the volumes and rates required for the economy of scale.

Another difficulty is that there seems to be no existing data on stover conversion using the process outlined here other than simulated models, and although it is believed to be effective, confirmation of several factors needs to be obtained. These include: (1) effectiveness of cellulase enzyme and its production on stover substrate, (2) viscosity and physical characteristics/behavior of slurry at several critical process points – necessary for proper equipment selection, (3) evaluation of alternatives to the ion exchange and overliming processes for removal of acetic acid and other substances toxic to fermentation, (4) alternative pretreatment reactor configurations, i.e. batch, (5) characterization and market development for the lignin waste or possibly re-evaluation of on-site combustion with electricity generation.

The evaluation of possible cellulase sources (on-site reference model produced, on-site PureVision produced, or purchased cellulase) strongly suggests that on-site cellulase production is not simply a resourceful idea, but a requirement. In addition, on-site cellulase production with the PureVision technology can mean significant savings in annual cost, even over the reference case model of cellulase production.

Each of the above issues, taken individually, has significant capital and operating repercussions. Combined, they have a considerable impact on the overall economic feasibility of the facility. Further discussion on these issues can be found in section 12 of this report.

2 INTRODUCTION

The biofuels program at the National Renewable Energy Laboratory (NREL), under guidance from the Department of Energy (DOE) Office of Fuels Development (OFD), is working to facilitate the commercialization of lignocellulosic biomass, i.e. corn fiber, corn stalks and wood to ethanol for use as a transportation fuel. OFD's ultimate vision is the large-scale production of ethanol from biomass to serve the nation's transportation needs.

To make this vision a reality, OFD supports research of process technologies, feasibility studies, and related commercialization activities by national laboratories, universities, private industry, research foundations and other government entities. In addition to technical achievement, substantial market development must also occur with expectation that industry leaders will emerge as the route to commercialization is clarified.

2.1 BUILDING THE BRIDGE

OFD recognizes the leadership potential of the existing grain (corn) processing industry. Their resources and experience provide the grain processing industry with the ability to lead commercialization of biomass to sugars and ethanol. The grain processing industry is the largest contributor to current ethanol and sugar production.

Recent feasibility studies for the production of sugars and ethanol from biomass at green-field sites have shown that capital expenditures contribute a large fraction of the cost, and must be reduced if the conversion process is to be economically viable in the near term. Adding to an existing ethanol plant or other site with compatible processes may reduce capital and operating cost. Roads, utilities, other process and operational infrastructure may be able to support increased operations and reduce the cost of sugar and ethanol production. Increased process utilization may also be possible.

2.2 PROCESS TECHNOLOGY

NREL supplied a detailed description of a corn stover to ethanol process including process flow diagrams, material balance, equipment descriptions and costs[11]. The NREL process uses simultaneous saccharification and co-fermentation (SSCF) and the design is based on a 2000 dry metric ton per day corn stover rate. The published design noted in the "References" section as (11) is a general lignocellulosic design based on yellow poplar. For mass balance purposes, NREL produced an identical model reflecting the use of corn stover as feedstock and issued it as a "Technical Memorandum." This was used to develop the 45% scale mass balance and is considered the "reference model" throughout this report.

The process selected for this evaluation uses on-site production of cellulase via a proprietary process and separate saccharifcation (hydrolysis) and co-fermentation (SHCF). The cellulase production is based on laboratory findings developed by PureVision Technology, Inc. (hereafter PureVision).

A plant feed rate of 900 dry metric tons/day of corn stover was selected based on readily available corn stover in the vicinity of the existing York, NE plant (Appendix 1). This rate is 45 % of the reference model.

2.3 BIOMASS FEEDSTOCKS

Biomass feed stocks comprise one of the largest sustainable resources on earth. They are produced in quantity from agricultural and forestry activities and are largely considered to be residue and waste. Locating a biomass conversion facility close to the feedstock can minimize the cost of transporting the materials. Facilities that produce biomass-derived products and are in the area of crop production (such as corn-to-ethanol facilities) have ready access to low-cost biomass feedstocks.

Grain processing sites are located near grain and agricultural residues. Corn stover is the single largest agricultural residue. Most grasses, hays and straws have cellular structures similar to corn stover, so a conversion technology that will work with corn stover will be likely to work with these other potential feed stocks.

Processing starch from corn to ethanol in a dry mill produces spent grain, which is sold for animal feed (distillers dry grains - DDG). With recent decline in the market and value of animal feed, dry mill fuel ethanol facilities need to find other methods to ensure economic health aside from the high protein and fiber feed DDG. One possible method is to use lower cost feedstock. Corn stover fiber left in the fields as agricultural waste can provide just such a feedstock for fuel ethanol production.

2.4 CELLULASE ENZYMES

The cost of cellulase enzymes is important to the commercial viability of a biomass conversion facility. In 1997 NREL performed an assessment of cellulase enzymes utilizing worldwide industry and academia input. The consensus position captured by the assessment showed cellulase enzyme costs could be lowered 5 to 10 fold by using proven biotechnology tools.

PureVision has been pursing this goal for several years, most particularly for the conversion of waste paper to glucose. Their findings already show improvement over more conventional cellulase production processes.

With the proprietary Pure Vision enzyme production process, a biomass-to-ethanol facility can produce enzyme that has a specific activity (effectiveness) of 800 FPU/g protein as opposed to the current lignocellulosic model of 600 FPU/g protein. If the same dose of cellulase (15 FPU/g cellulose) is used in enzymatic hydrolysis as is used with the reference model, the result is a decrease in feedstock flow to enzyme production of 25%. The cellulose that would ordinarily be consumed in enzyme production is now available for hydrolysis to sugar and further conversion to ethanol. However, as mentioned earlier, the significantly reduced hydrolysis time (~57%) results in a lower hydrolysis efficiency (84% as opposed to 88%) than the NREL reference model.

Another benefit of the enzyme's higher effectiveness is that the time required for hydrolysis is reduced from 48 hours to 18 hours (although a more conservative 24 hour enzymatic hydrolysis is used in this study). These two benefits result in a decrease in capital cost for enzyme production by reducing the number and scale of equipment items required.

The Pure Vision enzyme benefits come with no additional increase in equipment items, chemicals, or operating requirements other than the addition of a proprietary "very small amount" of a "low molecular weight enhancement factor." The enzyme is also produced with the same yields and protein productivity rates as the reference model[7] (see Table 6.2.3.B).

3 PURPOSE

The Corn Stover to Ethanol Process Evaluation Project explores the business potential of producing fuel ethanol from corn stover. Evaluation of the commercialization possibilities is based on co-location at, and shared infrastructure with, the existing High Plains Corporation (hereafter High Plains) corn-to-ethanol plant in York, Nebraska.

NREL has defined a benchmark process technology, including process flow diagrams, material and energy balances, required equipment, and the performance of cellulase hydrolysis and subsequent fermentations. NREL has also provided an estimate of the ethanol production costs for a new stand-alone facility built to the benchmark specifications. The NREL "reference lignocellulosic plant" is sized for 2000 bone-dry metric tons per day of corn stover feedstock and produces approximately 58.5 million gallons per year fuel ethanol at a total production cost of $1.30 per gallon.

The purpose of this evaluation is to develop and identify an alternative addition to the existing High Plains Corp., York, NE grain-to-ethanol facility to enhance overall economics of fuel ethanol production. This is to be accomplished by applying the reference lignocellulosic model developed by NREL and producing a process design, material balance, and capital and operating cost for the co-located facility. Modifications to the reference model include recent advances in the production and effectiveness of cellulase enzyme by PureVision Inc. Unlike the NREL reference design, the plant studied here uses separate hydrolysis and co-fermentation as well.

4 SCOPE

The overall scope of this study is to investigate the addition of a facility, not vastly dissimilar to the NREL reference-type lignocellulosic plant, to the existing High Plains facility, determine the approximate optimum production capacity of the added plant, and then evaluate the resulting production costs for the additional ethanol. The infrastructure and capacity resources of High Plains are utilized to reduce the capital and operating expenses of additional ethanol production.

Stover is pre-treated with dilute sulfuric acid, hydrolyzed using cellulase produced on-site via Pure Vision enzyme production technology, then co-fermented for the production of alcohol. Merrick has produced material balances and updated the NREL process flow diagrams and equipment lists. Merrick has also compiled a new project Pro Forma for the co-located plant, identified parameters that most significantly impact production costs, and performed sensitivity analyses on those parameters. Additional sensitivity analysis will be performed to assess the economic effect of obtaining required cellulase enzyme from various sources. Merrick will also define what effect co-location with the existing York facility has on the economics of a lignocellulose-to-ethanol facility.

5 FEEDSTOCK DESCRIPTION

5.1 CORN STOVER

The area surrounding the High Plains Corp., York, Nebraska, corn-to-ethanol plant is a prime agricultural area for the growing of corn. This study is based on the collection of stover from the five counties adjacent to and including York County. A circle of collection centered about the plant was not used, as highway access and stover yield data indicated a more practical method using the county boundaries. The maximum effective transportation distance is approximately 70 miles.

There are many variables in the corn stover collection system which could affect the quantity of stover available for processing. The following factors were used in sizing the plant to be evaluated:

- 60 wt.% of the corn stover can be collected from the fields in an economical and practical manner.
- 50% of the corn producers in the area will participate in the collection program.
- Available stover ranges from 2.0 short tons per acre to 3.7 short tons per acre.

A conservative decision was taken to use 1323 short wet tons (32.0% moisture) per day of collected stover (equivalent to 900 bone dry metric tons per day of stover) resulting in approximately 25.7 million gallons per year of ethanol production. Appendix 1 contains the detailed information regarding feedstock supply assessment. This information makes various assumptions and is of a different design basis than the facility modeled in terms of ethanol produced and the sizing of the facility; however, the "Total Tons for Biomass Conversion" provides the average to be used in sizing this facility.

The proven method of collection is to rake stover, which is left in the field either scattered or as a windrow by the corn harvesting combines. The stover is baled at the site in large cylindrical bales. A well-made bale is 1.52 meters tall and 1.78 meters in diameter (5 foot tall, 70 inches in diameter) and weighs about 544.3 kg (1200 pounds). Bales bound by a triple wrap of plastic netting have proven to be more economically attractive than twine bound bales as there is less loss during highway travel and better retention of the bale shape during storage[10]. Bales will be transported using trailers pulled by highway legal tractors or by trucks. Regional collection and storage facilities are felt to be more practical than storing bales at individual farms although this subject requires further inquiry. Bale storage at the plant will be the equivalent of four days of plant feed. The harvest of stover is believed to last a maximum of 120 days. Please see Appendix 2 "Trip Report April 1 and 2"for more detailed information regarding harvest and transport of stover.

5.1.1 Total Sugar/Lignin/Ash

The composition of corn stover in Table 5.1.1 was taken from NREL Technical Memorandum "Modified Process Model Results for a Feedstock Composition Reflecting Corn Stover", April 26, 1999 which cites *Renewable Energy*, October, 1997, "Bioethanol Production: Status and Prospects", J. McMillan[2]. This composition is used as the basis of this study:

Table 5.1.1: CORN STOVER COMPOSITION

Component	Weight % Dry Basis
Cellulose	45.39
Xylan	23.86
Arabinan	2.00
Mannan	0.00
Galactan	1.11
Acetate	2.11
Lignin	18.53
Ash	7.00
Total	**100.00**

Note: The compositions here (which are the design basis of this study) are different than those assumed in Appendix 1 page 4.

5.1.2 Estimate of Cost

The High Plains Corporation working with privately held data has estimated the cost of corn stover delivered to the plant site as less than $35.00 per dry short ton (see Appendix 1). This figure is valid only after regional collection/holding centers are established, harvesting machinery is available and some other start-up costs are paid off. This is a reasonable maximum cost in the third year of corn stover collection.

Further, the cost of delivered stover will likely fall to as little as $20.00 per delivered dry long ton when its collection and storage are well established (presuming that a competitive corn stover market does not develop in the area).

5.2 DISTILLERS GRAIN

Distillers grain was considered for feed along with the corn stover but was not included. **The distillers grain was not included because of its high value as an animal feed.** Distillers grain is valued for animal feed based on its protein content and could, therefore, pass through saccharification without significant loss of value. However, the solids from saccharification of the distillers grain would need to be kept separate from the solids from the corn stover as the corn stover solids have little or no value as animal feed and are intended to be sold as a fuel. In essence, this means that distillers grain would require separate processing facilities. In addition to this, the mixing of distillers grain with the genetically modified *Zymomonas mobilis* used for co-fermentation greatly decreases its marketability. For these reasons, processing distillers grain is not justified for the small amount (350 tons/day) and high value ($60 to $90 per ton) of the distillers grain in the current local market. See Appendix 1 for details regarding the assessment of the use of distillers grain as lignocellulosic feedstock for increased ethanol production.

6 FACILITY DESCRIPTION

6.1 HIGH PLAINS CORN TO ETHANOL PLANT

6.1.1 Facility Production Capacity

The existing grain to ethanol plant at York uses a dry mill process, consuming 351,081 wet metric tons (387,000 long tons) per year of corn and milo to produce 37.5 million gallons per year of ethanol.

Feed grain is delivered to the plant via truck. Of the 37.5 million-gallons/year ethanol production capability, up to 12 million gallons can be further purified, in a separate distillation section, to industrial grade ethanol.

High Plains has the capability to store up to 7 days of grain feedstock in 4 silos. They use a single day bin to feed 3 hammer mills that grind up to 45,000 bushels/day. The mills have dust control cyclones and a bag house with pulsejet cleaning of the bags. Recovered dust is added to the ground feed and travels with it. Each hammer mill has an outlet screen to control particle size of the grind. The grain is ground to coarse flour. The flour is conveyed in an elevated conveyor system to the slurry tank. Following milling, recycle water from multiple sources (backset), ammonia for pH control and an α-amylase are added in the Slurry Tank which operates at about 65.6°C (150°F). Next, this slurry is pumped and mixed with steam in a Hydroheater to bring the temperature to 107.2-121.1°C (225-250°F). The Hydroheater discharges to the bottom of the Cook Tube, which has a 20-minute residence time, and up-flows into the flash tank. The slurry is flash cooled at a slight vacuum (the source of the vacuum is the Rectifier Tower overhead vacuum system) to a temperature of approximately 87.8°C (190°F).

Additional α–amylase is added to the slurry, which is then held in Liquifaction Tanks (plug flow horizontal tanks having three mixed chambers in each tank) for approximately two hours. The liquefied "mash" then flows into a second vacuum flash cooling vessel (vacuum is generated through condensing and vacuum pumps) to lower the temperature to 62.8°C (145°F), before being fed into the saccharification tank. In saccharification, sulfuric acid is used to lower the pH to the desired level for enzyme activity and glucoamylase is added to begin converting the starch into sugar (20 minute hold time). A side stream of sugar is taken for the production of yeast in a separate vessel. Yeast is propagated for 5 hours before being pitched into the filling fermentor. Each fermentor receives 2 to 3 pitches of propagating yeast. The mash flows from the saccharification tank through spiral heat exchangers (scrolls) to reduce the temperature from 62.8°C (145°F) to 29.4°C (85°F). There are 9 spiral exchangers - three parallel trains having three exchangers in series in each train, all feeding the selected fermentor.

The fermentors are 15.24 m (50 feet) in diameter by 15.24 m (50 feet) tall and have a 2,460,518 L (650,000 gal) working volume in each. Fermentors go through a 40-hour cycle - 17 hours to fill, 17 hours residence and 6 hours to empty and clean in place. During filling, at 10% full and 50% full, yeast is added from the yeast propagators. Fermentors have 4 loops of cooling coils in each. A batch normally is fermented to 13% alcohol. Carbon dioxide evolved from the fermentors is scrubbed (counter-current) with water to remove particulates and soluble (volatile organic) emissions and then vented to the atmosphere or transferred via pipeline to a carbon dioxide refiner (on site customer).

There are three fermentors. A fourth, 2,725,496 L (720,000-gal) vessel functions as a surge vessel between fermentation and distillation. This surge vessel is called the Beer Well.

Distillation is conventional, having a beer or stripping column with water and alcohol overhead and solids and water out the bottom. Stripper overhead feeds the middle of the rectifier column, containing 13 stripping trays below and 40 concentrating trays above.

Rectifier column overhead goes to mol sieve dryers (3 operating and one on stand-by) each having an 8-minute cycle time on duty and 8 minutes regenerating (water purge).

This plant also has an industrial alcohol distillation unit, which produces higher purity alcohol than required for fuels. It is fed with a side draw taken from the second or third tray in the top of the rectifier. Water is added as a wash/stripping agent and the alcohol is re-distilled to 190 proof grain neutral spirits. A future molecular sieve dryer is planned.

Slurry from the beer column bottoms is fed to Sharples horizontal decanter type centrifuges where the soluble portion is separated from the insoluble fiber. The soluble stream is fed to evaporators, which concentrate the stream to syrup. The syrup is then blended with the solids from the centrifuges to produce the distillers grain. Distillers grain with solubles (DGS) is often sold wet to local feed lots. If the distillers grain must be dried, the drying is done in gas fired rotary dryers (kiln type).

6.1.2 Site Description

York, Nebraska is located half way between Lincoln and Grand Island or approximately 160.9 km (100 miles) west of Omaha on Interstate 80. The plant is located in a rural setting, 8-10 km (5 to 6 miles) from the town of York. There is excellent highway and rail access to the site.

6.1.3 Infrastructure Description

The plant employs about 55 people. There are approximately 33 people in operations with the remainder in administration and maintenance.

The existing facilities include a laboratory, shops and warehouse, office, parking areas, security, communications, road and rail access and other features common to stand-alone industrial facilities.

A Johnson – Yokogawa (Yokogawa Industrial Automation) distributive control system, provides process automation for micro processing and analog input/output control. It can be expanded to handle the new processing facilities.

6.1.4 Utilities

Two cooling towers provide heat dissipation for the processing. One tower circulates approximately 64,352 L (17,000 gal) per minute and a smaller tower circulates approximately 37,854 L (10,000 gal) per minute of water. Both towers are designed for 10°F cooling. The cooling water distribution system is designed for flexible operations, e.g. cooling the Industrial Alcohol Distillation and Fermentation with the small tower and cooling the remainder of the processes with the large tower. Makeup water is from wells located on the property. Well water is softened and treated with reverse osmosis (primarily for boiler water feed) prior to use. Any excess treated water is used for cooling water make-up. Blow-down from the small tower is used for make-up water to the larger tower. Any additional make-up water required is from untreated (high hardness) well water. Cooling tower blowdown is discharged to a lift station, combined with pre-treated wastewater and pumped directly to the city sewers. Total well water usage is approximately 75.7 L (20 gal)/bushel of feed. A majority of water requirements are for boiler water and cooling tower makeup.

Chilled water is provided by two, 900 HP, motor driven, York self-contained, mechanical refrigeration (Freon) machines. They are only needed for control of the exothermic fermentation in the summer. Normal chilled water temperature is 15.6°C (60°F).

The mills, conveyors, mixers, fans and many centrifugal pumps are direct driven by electric motors. The centrifuges and some centrifugal pumps are driven by variable speed electric motors. Total power consumption is approximately 1.3 KW per gallon produced with a peak demand of 5800 KW.

The plant consumes approximately 5500 MMBtu/day of natural gas mainly in the boilers with approximately 10 - 20% used in the distillers grain dryers. Total steam available from the two boilers is 200,000 pounds per hour at 150 PSIG. Typical steam usage is 130,000 PPH with the Industrial Distillation System in operation.

6.1.5 Recycle Water

Several streams feed water to the Recycle Tank. 25% of the evaporator condensate (remaining goes to wastewater treatment), and all of the Rectifier Column Bottoms go to this tank. From the Recycle Tank, water is fed into the

Slurry Tank for mixing with the flour (ground grain) and added to the Saccharification Tank for solids control, thus reducing wastewater volumes and minimizing make up water requirements.

6.1.6 Waste Disposal

Condensate from the evaporators having 1500 to 2000 mg/liter COD is feed to anaerobic digestion. Anaerobic digestion (methanators) consists of 4 – 113,562 L (30,000 gal) fiberglass vessels, arranged in parallel, and providing 6 hours of residence time. Methanators are sized for 2 gal/sq.ft./min. of liquid flow. They are designed for 90 % COD reduction to less than 200 mg/liter COD and have only 3% sludge in the treated water. They operate at 35^{O}C (95^{O}F) and use micronutrients for organism health and caustic for calcium requirements and pH control.

Methanator liquid output goes to aeration ponds for additional treatment, then a clarifier for solids removal, and then combines with untreated waste (boiler and cooling tower blowdown) to be pumped into the city sewage system. The clarifier is a conventional circular, cone bottomed type with scrapers on the cone. Activated and other settled sludge is pumped from the bottom and returned to the aeration pond. Excess sludge may be wasted to a retention pond for 30-day aeration before being land applied.

6.1.7 Roads and Railways

The DDG and ethanol products can be shipped via truck or rail. The corn and milo feedstock is delivered by truck.

The plant is located on Highway 34 approximately 4.8 km (3 miles) east of the interchange with Highway 81 and 11.3^{O}C (7 miles) north of Interstate 80. Road access is felt to be adequate for the delivery of corn stover.

The plant has a 75-car capacity rail siding with dual spurs connected directly to a BN-SF main line. An on-site car mover is utilized and BN-SF provides up to two switches per day, 6-days per week

6.2 CORN STOVER PROCESSING REQUIREMENTS

6.2.1 Feed Receiving and Handling

A. Overview

Currently, stover is harvested from the field and baled from the ground in large round bales then transported to processing facilities by flat bed trailer. When considering the bulk handling issues such as bale damage, removing bale wrappings, field and other debris, and the large volumes of material, we have determined that an alternative method for feedstock harvesting and handling needs to be studied further. Conversations with Iron Horse Custom Harvesting indicate that this point has been recognized and they have devised for study, such alternative methods[10]. Further discussion of this can be found in the "For Further Study" section of this report.

Corn stover is delivered to the High Plains York Ethanol production facility on trailers pulled by high-speed tractors. The trailers are weighed and then unloaded onto a concrete pad. Loaders then either stack or move the stover to feedstock conveyors, which convey the bales into a processing unit. The processing unit debales and shreds the stover. The shredded stover is then conveyed to a concrete bunker. A loader pushes shredded stover from piles in the bunker to a pretreatment feed conveyor. This conveyor feeds the pre-treatment reactor.

Although there is no washing of stover designed into this facility, NREL experience with the Process Development Unit (1 ton/day pilot plant,PDU) shows that the feedstock needs to be quite clean to reduce *lactobacillus* contamination and to decrease wear on the pre-treatment reactor[8]. However, in washing stover there is a potentially significant loss of feedstock to water that will need to be sent to wastewater treatment. This study assumes that the bales are quite clean of soil and on site they are stored on a concrete pad where large amounts of soil can be manually removed with hoses if necessary. This is similar to the original process design of the NREL PDU for municipal solid waste[1]. The bale breakers have the ability to remove tramp metal debris.

The bales have an assumed moisture content of 32%. The feed stream of shredded stover into the pre-treatment reactor needs to be 48% moisture. The above mentioned washing, water mist added during the shredding process to reduce dust and fire hazard, and climate conditions experienced by shredded stover in the shred bunker are assumed to bring the moisture to the 48% level.

B. Design Basis

Process Flow Diagram -P101-A101 (all PFD's are in Appendix 4)
Corn stover is feed into the pretreatment reactor M-202 at a rate of 71,977 kg/hr at 48% moisture. Operation of the reactor is for 24 hours each day, 350 days each year. This requires the delivery of 1,654 bales per day at an average bale

dry weight of 544.3 kg (1200 pounds). Each truck delivery capacity is 17 bales, with each bale measuring 1.52 meters tall and 1.78 meters in diameter (5 feet long and 70 inches in diameter). Average water content[10] is 32%, with a dry mass bale composition of 82% stalk and 18% cob[13].

Due to the "wide load" status of the delivery trailers and possible state highway laws[3,10], it is assumed that delivery will be 5 days per week. Therefore, design capacity for bale receiving and processing is 2,315 bales per day. This requires 136 deliveries per day using two truck scales (M-101A/B, not including the scales that currently exist at the York facility). Trucks can be weighed, sampled for moisture, and unloaded in less than 10 minutes[9]. Bales are stacked in rows, two bales high and transported to the bale processing feed conveyor as needed by 6 forklifts or loaders at a rate of 30 minutes per 17 bales (truck load). Unloading, stacking, and transport is all done on a 23,226m^2 x 22.9 cm thick (250,000 ft^2 x 9") concrete pad (M-102). The pad has surface area for 7440 bales (four and a half day feed) at 2.79 m^2 (30 ft^2) per bale. The bales are stacked two rows high and the pad has an area for vehicle maneuvering equal to the bale storage area with an additional 10% area for drainage. Storm drainage is collected in pond M-108 from which flow to waste water treatment is metered at 39,407 kg/hr. Specifications and calculations for feed stock handling can be found in Table 6.2.1.

Table 6.2.1: Feedstock Handling Calculations and Assumptions

bales (each)	
weight dry (#)	1200
weight wet (#)	1600
% solids	68.0%
length (in)	60
diameter (in)	70
% stalk (w/w)	82%
% cob (w/w)	18%

trucks (each)	
# / delivery (dry)	20,400
bales / delivery	17
Delivery days	350

plant run time	
days / year	350
% of year	96%
hrs / day	24

feed rate		
kg / hr wet (101 PFD-A101)	71,977	
% solids	52.1%	
# / hr (dry)	82,662	37,489kg/hr (dry)
ton / hr (dry short ton)	41.3	37.5metric tons
ton / day (dry short ton)	992	900metric tons
bales / day	1,653	
ton / day (wet short ton)	1,323	1,200metric tons
ton / year (wet short ton)	347,182	
bales / year	578,637	
feedstock spec. size (in)	3/4 x 5/8 x 1/8	

deliveries		
bales / day	1,653	2,315design capacity (7/5)
bales / hr	138	193 "
trucks / day	97	136 "
weigh time / truck (min.)	10	10 "
delivery hrs / day	12.0	12.0 "
deliveries / scale / day	72	72 "
number of scales required	1.4	1.9 "

receiving and processing		
bale receiving pad (ft^2)	250,000	
dimensions (ft)	500	
forklift time per truck (17 bales) (min.)	30	
forklifts/loaders	5	6 "
bale processing (wet short ton / hr)(12hr/day)	122	170 "

Storage	
days of storage	3
shredded density (wet #/ft^3)	15
bunker volume (wet short tons)	4,376
Bunker volume (ft^3)	583,499
(200x100x30) =	600,000

waste water run-off calcs.	
bale receiving pad (ft^2)	250,000
precipitation (in/hr)	2
storm hours/wk	5.6
run-off to WW treatment (gal/hr)	311,688
storm run-off (gal/wk)	1,747,767
flow through WW treatment (gal/hr)	10,403
kg/hr to WWT	39,407
Holding pond (one week)(ft^3)	233,643
dimensions 200 x 150 x 8ft (ft^3)	240,000

loader fuel	
Loaders	7
loader hrs / day	87.6
fuel usage (gal/hr)	8
gal/day	701
kg/day	1907

Bales are received and processed for 12 hours each day. As the six front-end forklifts/loaders stack and transport bales to the bale feed conveyors (C-101), operators cut and remove the plastic netting using hooks. Netting is added to the gypsum waste produced in area 200 (overliming) and is insignificant in weight in relation to the gypsum. These materials are landfilled and the cost of netting disposal is included in the gypsum disposal cost.

The unwrapped bales are then conveyed to the bale breaker M-104 and the primary and secondary shredders M-105 and M-106. Shredded stover is then conveyed by the radial conveyor C-102 to a shred bunker (M-107) that is 61m long x 30.5m wide x 9m tall (200ft long x 100 ft wide x 30ft tall) and has a three-day capacity of 16,990 m^3 (600,000 ft^3). The bottom of the shred bunker has a screw conveyor C-109, which is assisted by a loader to assure continuous feeding of the pre-treatment reactor.

C. Cost Estimation

Cost estimation for the truck scales and storm runoff pond came from recent Merrick experience, as did the design and cost estimation of the receiving pad and the shred bunker. Vender quotes from American Pulverizer are the design and cost basis for the hammer mills and associated conveyors. The wire mesh bale conveyor was vender quoted by Conveying Industries. The radial stacker was designed and costed by SESCO conveyors and engineering. The three bale breakers are ADB Series II from Karl Schmidt and Associates, Inc. handling 700 ton per day each. The pre-treatment feed conveyor and loaders were scaled from the lignocellulosic reference model produced by NREL.

6.2.2 Feed Pretreatment

A. Overview

Shredded corn stover is conveyed to the pretreatment reactor where it is hydrolyzed with high temperature, pressure and dilute sulfuric acid. The hemicellulose portion is broken down to simple sugars with xylose being the primary product. In addition, some arabinose is released. This process results in the production of some acetic acid, furfural, and hydroxymethyl furfural (HMF) as by products. The lignin-cellulose complex is also broken down resulting in some glucose, mannose, and galactose from the cellulose, but of primary interest is the exposing of the cellulose for the following enzymatic hydrolysis to glucose.

The pretreated stover is then flash cooled resulting in a significant reduction in water content as well as a reduction in furfural and acetic acid. Due to the toxic nature of the remaining acetic acid, furfural, and HMF to enzymatic hydrolysis and fermentation the solids (primarily lignin and cellulose) are separated from the liquid (xylose, soluble sugars, acetic acid, water, furfural, and HMF) so that this liquid can be detoxified.

Detoxification is done with continuous ion exchange followed by an "overliming" process. There is currently research underway at NREL[15,16], which is trying to understand the importance of overliming to the prevention of toxic conditions in fermentation. Gypsum is produced as a waste product from this area. The pH is also adjusted to 4.5 in preparation for enzymatic hydrolysis. The detoxified liquid is re-mixed with the cellulose and lignin solids and distributed to fermentation seed production, cellulase seed production, cellulase production and enzymatic hydrolysis. This liquids and cellulose/lignin solids mixture will be called the liquor.

B. Design Basis

Process Flow Diagram -P100-A201

The corn stover from the screw conveyor C-109 is warmed with direct injection low-pressure steam in M-202 to 100°C. Condensate is mixed with sulfuric acid and added to the warmed stover in the impregnator portion of the reactor to a total sulfuric acid concentration of 0.5% of the total amount of steam, condensate, and stover. High-pressure steam (265 °C) is used to bring the reactor to 190 °C for 10 minutes (Table 6.2.2.A).

Table 6.2.2.A: Pretreatment Reactor Conditions

Acid Concentration	0.5%
Residence Time	10 minutes
Temperature	190 °C
Solids in Reactor	22%
Reactor Pressure	12.2 atm

Pretreatment reactions and conversions occurring in the hydrolyzer are from NREL[11]. These are contained in Table 6.2.2.B.

Table 6.2.2.B: Pretreated Hydrolyzer Reactions and Conversions

Reaction	Conversion
$(Cellulose)_n$ + n H_2O → n Glucose	Cellulose 0.065
$(Cellulose)_n$ + m H_2O → m Glucose Olig	Cellulose 0.007
$(Cellulose)_n$ + n H_2O → ½ n Cellobiose	Cellulose 0.007
$(Xylan)_n$ + n H_2O → n Xylose	Xylan 0.75
$(Xylan)_n$ + m H_2O → m Xylose Olig	Xylan 0.05
$(Xylan)_n$ + → n Furfural + 2n H_2O	Xylan 0.10
$(Xylan)_n$ + n H_2O → (Tar)n	Xylan 0.05
$(Mannan)_n$ + n H_2O → n Mannose	Mannan 0.75
$(Mannan)_n$ + m H_2O → m Mannose Olig	Mannan 0.05
$(Mannan)_n$ + → n HMF + 2n H_2O	Mannan 0.15
$(Galactan)_n$ + n H_2O → n Galactose	Galactan 0.75
$(Galactan)_n$ + m H_2O → m Galactose Olig	Galactan 0.05
$(Galactan)_n$ + → n HMF + 2n H_2O	Galactan 0.15
$(Arabinan)_n$ + n H_2O → n Arabinose	Arabinan 0.75
$(Arabinan)_n$ + m H_2O → m Arabinose Olig	Arabinan 0.05
$(Arabinan)_n$ + → n Furfural + 2n H_2O	Arabinan 0.10
$(Arabinan)_n$ + n H_2O → (Tar)n	Arabinan 0.05
Acetate → Acetic Acid	Acetate 1.00
N Furfural + 3 n H_2O → (Tar)n	Furfural 1.00
N HMF + 3 n H_2O → 1.2 (Tar)n	HMF 1.00

Note: These reactions are modeled as occurring simultaneously. Therefore, products of one reaction, e.g., furfural, are not considered a reactant in another reaction. Degradation of xylan all the way to tar is accounted for as a reaction of xylan to tar. Degradation of furfural considers the furfural entering the reactor in the recycle water.

The pretreated stover liquor is flash cooled for 15 minutes in T-203 to atmospheric pressure where 6.3% of acetic acid and 61% of furfural and HMF are removed. The 190 °C flash vapor is used to preheat the beer to ~95 °C in H-201 on its way to the beer stripping column. The condensed flash vapor is then sent to waste water treatment at ~99 °C (NREL[11]).

Process Flow Diagram -P100-A202

The solid and liquid portions of the pretreated slurry from T-203 are separated with a washing belt filter press S-202 to produce a solids portion of 40% insoluble solids and a liquid portion. The liquid portion and the filter rinse water are pumped with P-227 to ion exchange after being cooled to 40 °C in H-200 with cooling water. Approximately 88% of the acetic acid and 100% of the sulfuric acid are removed in the continuous ion exchange unit (S-221), which is regenerated with ammonia at 1.1 normal per normal of ions. Further discussion of its treatment (overliming) is to follow.

The solid portion (lignin and cellulose) is transferred to T-232 via a screw conveyor C-202. Here the solids and detoxified liquid returning from overliming are mixed for 15 minutes with 2hp/1000gal. The pretreated and detoxified stover slurry is then pumped with a 700 gpm Discflo pump (P-224) to hydrolysis (86.7%), fermentation seed production (9.5%), cellulase seed production (0.2%), and cellulase production (3.6%). Table 6.2.2.C illustrates a comparison between the reference model flow rates and the co-located flow rates.

Table 6.2.2.C: Flow Rate Comparison with the Reference Model

	York Co-located Study Case		NREL Ref. Model with 45% Scaledown	
Flow from mix tank (g/hr)	167,795,100	100.0%	167,795,100	100.0%
Flow to hydrolysis (g/hr)	145,536,045	86.7%	143,425,350	85.5%
Flow to seed production (g/hr)	15,936,745	9.5%	15,936,750	9.5%
Flow to cellulase seed production (g/hr)	315,559	0.2%	420,750	0.3%
Flow to cellulase production (g/hr)	6,006,751	3.6%	8,009,100	4.8%
total outflow	167,795,100	100.0%	167,791,950	100.0%
cellulose to be hydolyzed (g/hr)	15,014,766		14,818,500	

Process Flow Diagram -P100-A203

The liquid portion from ion exchange is overlimed by reacidification to pH 2 with addition of sulfuric acid using in-line mixer A-235. This is mixed with lime to pH 10 in T-209 for one hour with steam injection to 50 °C. Mixing is accomplished with 0.5 hp/1000 gal. The pH is then adjusted to 4.5 in T-224 with residence time of 4 hours. Again, mixing is accomplished with 0.5 hp/1000 gal. The liquid and resulting gypsum are separated with 99.5% gypsum removal (containing 20% liquids) by a hydrocyclone and rotary drum in series. The detoxified and conditioned liquid is then recombined in T-232 as described above.

C. Cost Estimation

Several equipment items in area 200 had costs of greater than $100,000 per unit and so received new cost estimates. These include: T-224, which received a new price quote from Matrix Service, Inc.; and H-201 from Lawhorn Shell and Tube, Inc. Other "High cost" items such as the belt filter press (S-202), Sunds hydrolyser (M-202), and the ion exchange unit (S-221) did not receive new price quotes due to either their very large cost, or specialized nature. In either case it was felt by Merrick engineers that NREL had the best quotes available and these were used for scaling. The pump P-224 was changed to a Discflo because Merrick believed that this pump would better handle the high solids content of the liquor. All other equipment in area 200 was cost scaled from the NREL 2000 dry metric ton per day reference model. Gypsum waste disposal was discussed and considered by High Plains with York County waste disposal personnel. Considerations included land application and landfill. In this study the landfill option is the assumed disposal method at a cost of $33 per short ton. It was decided that the very large quantities of gypsum would be inappropriate for land application.

6.2.3 Enzyme Production

A. Overview

Enzyme is produced on site using the Pure Vision cellulase enzyme production process. This process includes two areas, the production of the enzyme producing seed (*Trichoderma reesei*) and the production of the enzyme itself. The seed production originates with the inoculation of one of three seed production trains with a culture grown in the laboratory. Each train has three vessels, which are sized to provide a 5% inoculation to the next vessel in the train. The third seed vessel then inoculates one of eleven cellulase production vessels. The substrate for these areas is detoxified stover. Enzyme produced here is then sent to enzymatic hydrolysis.

B. Design Basis

Process Flow Diagram -P100-A401 & A402

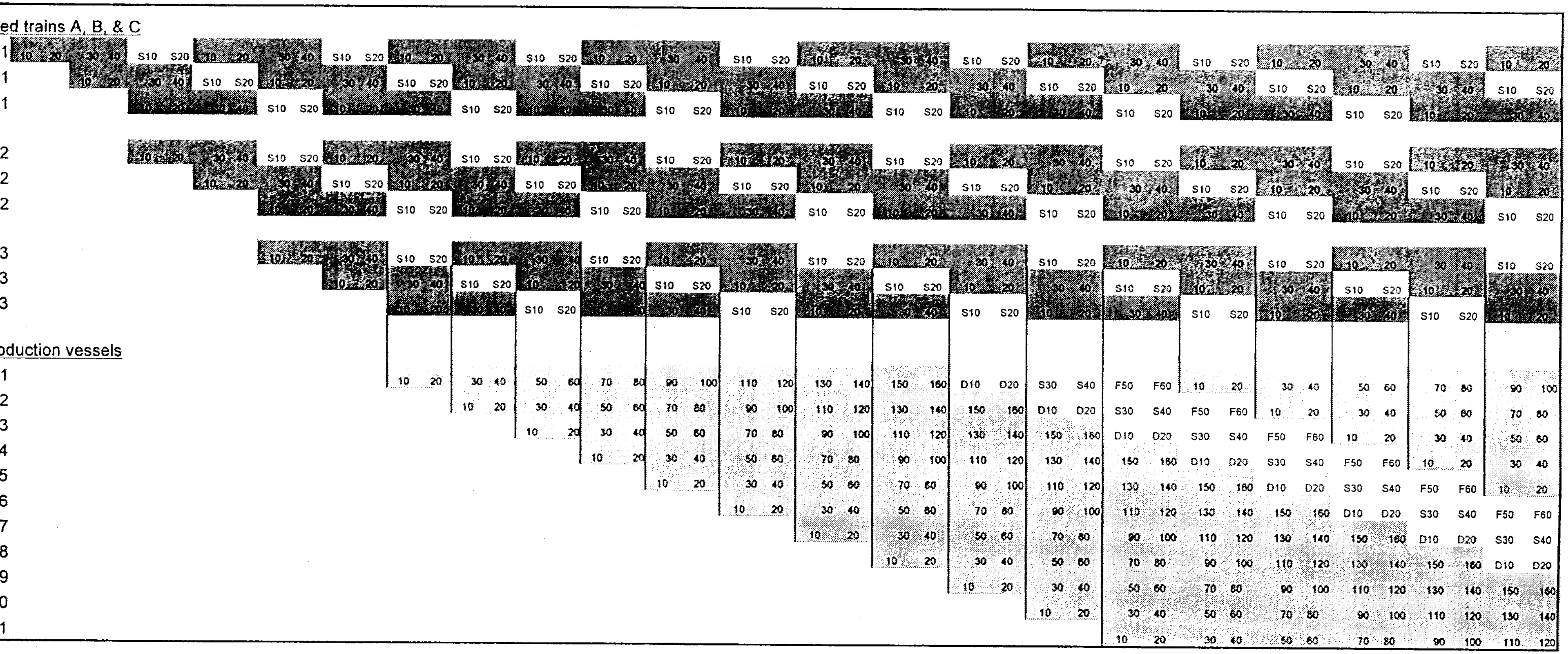
Although providing enzyme usage of 20 FPU and 35 FPU per gram of cellulose were initially considered, it was recommended by Jim Linden (CSU) and NREL that 15 FPU per gram cellulose would be more appropriate (Appendix 3). Using laboratory data provided by PureVision for their cellulase production technology, the feed rate to cellulase seed production is 0.2% of the pretreated slurry flow from T-232 and the feed rate to enzyme production is 3.6% of the pretreated liquor. Of the remaining 96.2% of pretreated slurry, 86.7% goes to hydrolysis and 9.5% goes to fermentation seed production.

The cellulase is produced in eleven 334,384 L (88,335gal) production vessels F-400, which are sparged at 0.413 VVM with sterilized air from compressor M-401. The vessels have a diameter to height ratio of 2:1, which as a result of preliminary study by NREL, is most effective to provide the estimated requirement of 30% dissolved oxygen[11]. This study also indicated that increased concentrations of oxygen above atmospheric would be investigated further. This may be very important to obtain the desired saturated oxygen without using a pressure vessel. Our vessel cost quote reflects atmospheric tanks although it may be necessary to use pressure vessels to increase the dissolved oxygen as per preliminary compressor calculations suggest (see M-401 calculations in Volume II). Eight of fermentors are in operation at any given time with the remaining fermentors cleaning, draining, or filling (see Table 6.2.3.A: Cellulase Production Schedule).

Cellulase production residence time was chosen to be 160 hrs in keeping with production time suggested by NREL and PureVision. At a flow rate to enzyme production of 9,533 kg/hr, it was decided that the same number of production vessels and configuration as NREL used in the lignocellulosic model (only smaller) was most appropriate to keep vessel fill time to a minimum and ensure a more accurate 160 hr average enzyme production time. Cellulase broth is pumped with P-400 to enzymatic hydrolysis and a small stream is also sent to fermentation seed production.

Enzyme is produced based on the parameters outlined in Table 6.2.3.B which also contains a comparison between cellulase production using the PureVision technology and the NREL reference model. Laboratory data from PureVision indicates that the specific activity for their cellulase is 800 FPU/ gram protein. The productivity and yield are the same as those stated by NREL to be 75 FPU/(L*hr) and 200 FPU/gram cellulose respectively[10].

ble 6.2.3.A: Cellulase Production Schedule



seed production vessel #1 (gal)	11	
seed production vessel #2 (gal)	221	draining (D), sterilizing (S), or filling (F)
seed production vessel #3 (gal)	4,417	
size of production vessels (gal)	88,335	

Table 6.2.3.B: Cellulase Production Parameters

	York with PureVision	NREL Reference Model (45% scale)
yield (FPU/(g cellulose+xylose))	200	200
productivity (FPU/(L*hr))	75	75
specific activity (FPU/g protein)	800	600
initial cellulose concentration	4%	4%
cellulase requirement (FPU/g cellulose)	15	15
enzyme production broth (kg/hr)	13,384	17,848
enzyme production broth (gal/hr)	3,533	4,712
production time / vessel (hr)	160	160
Size of production vessels (gal)	88,335	118,800
production vessel operating volume (gal)	70,668	94,240
number of vessels in operation (add 3 for cleaning)	8	8
% fill of vessel	80%	79%
Time to fill vessel (hr)	20	-
Temperature °C	28	28

The *Trichoderma reesei* "cellulase seed" culture is first grown in three trains of progressively larger tanks, each representing a 5% inoculation of the next larger size. With production vessels of 334,384 L (88,335gal), three smaller vessels; F-401 – 16,720 L (4,417 gal), F-402 – 836.6 L (221 gal), and F-403 – 41.6 L (11 gal) are used to provide sufficient seed culture for the production level (see Table 6.2.3.C). Residence time in each seed vessel is 40 hrs, which has been determined by NREL research to be enough time to grow cell mass[11]. Three trains of these three sized vessels allows for 20 hours of turn around time per train. It should be noted that if hydrolysate from the hydrolysers (T-307) in which cellulose has already been converted to glucose were used for cellulase seed production, the batch time could quite likely be reduced. This should not have a negative effect on the cellulase production to follow. However, in this study the seed is grown on cellulose slurry directly from the mix tank T-232.

Table 6.2.3.C: Cellulase Seed Production Parameters

% inoculation of production vessels	5.0%
volume of inoculant needed (gal/vessel)	3,533
inoculant needed every (hrs)	20.0
batch time for each seed production (hrs)	40
seed production vessel #1 (gal)	11
seed production vessel #2 (gal)	221
seed production vessel #3 (gal)	4,417
trains of vessels	3 - "A", "B", "C"

Cellulase production is conducted by filling a production vessel with detoxified stover liquor such that the slurry at working vessel volume will contain 4% cellulose after the addition of recycle water, corn steep liquor (1% of mixture volume), and nutrients. The pH is controlled with ammonia and foaming is controlled with corn oil (0.1% v/v of final mixture). Nutrient requirements for cellulase production from pretreated biomass are still under study at NREL, but are estimated to be those contained in Table 6.2.3.D (NREL[11]). In addition, "small amounts of low molecular weight proprietary enhancement factor" are required by the Pure Vision cellulase production technology.

All tanks and pumps are sterilized with hot caustic clean-in-place (CIP) solution between batches.

Table 6.2.3.D: Cellulase Production Nutrient Requirements

Component	Amount (g/L)
Urea	0.3
$FeSO_4 - 7\ H_2O$	.005
$ZnSO_4 - 7\ H_2O$	.0014
$(NH_4)2SO_4$	1.4
KH_2PO_4	2.0
$MgSO_4 - 7H_2O$	0.0016
$CaCl_4$	0.002
Tween 80	0.2

Note: The PureVision cellulase production requires a "very small amount of a low molecular enhancement factor" which is not included here.

C. Cost Estimation

The agitators for these tanks were quoted by Lightnin to provide 1.4hp/1000 gal. A new price quote was also obtained from Atlas Copco for the fermentor air compressor system, which at five compressors and one back-up, is less costly than the option of using two (and one back-up) of the lignocelluosic compressors to provide the required 38,809+ scfm of air. All other equipment was scaled from the NREL 2,000 metric ton / day lignocellulosic model.

6.2.4 Hydrolysis

A. Overview

Aside from the enzyme production technology from Pure Vision, the key difference between the current NREL lignocellulose-to-ethanol model and this plant design is the fact that we are performing hydrolysis and co-fermentation separately (SHCF) as opposed to simultaneously (SSCF). The justification for this is that each step has different optimum conditions (hydrolysis 50°C vs. co-fermentation 30°C). The common approach to cellulose-to-ethanol conversion is SSCF in which either an enzyme modified for optimum performance at lower temperatures, or an ethanologen modified for thermophilic conditions, or (more likely) a combination of the two, are used simultaneously. This compromise is an effort to avoid product inhibition of cellobiose and glucose in hydrolysis, which tends to be the rate-limiting step. That we are aware of, there are no such industrially used thermophilic ethanogens nor low temperature cellulases capable of this compromise. Therefore, we have decided to keep hydrolysis and fermentation separate to take advantage of the optimum conditions for each process.

In this process, the pretreated and detoxified corn stover slurry is first hydrolyzed in large mixing tanks with agitators and pump circulation. After 24 hours of mixed hydrolysis and saccharification, the (now thinner) slurry is pumped to co-fermentation (fermentation of pentosans and hexosans by a single organism). Conversion of cellulose to glucose was assumed to be 80% as recommended by PureVision[7] and confirmed by NREL researchers[8]. Please see Table 6.2.4.A for the conditions of enzymatic hydrolysis.

Table 6.2.4.A: Enzymatic Hydrolysis Conditions

% insoluble solids (21.4% total solids)	15.0%
temperature (°C)	50
time per slurry (hr)	24
flow per slurry (kg/hr)	157,136
% conversion cellulose to glucose (hydrolysis)	80.0%
% conversion (SSCF 48hr)	39.5%
(overall hydrolysis conversion)	84.0%

B. Design Basis

Process Flow Diagram -P101-A307

Detoxified stover slurry is pumped through H-308 to enzymatic hydrolysis in T-307 at a rate of 145,536 kg/hr (86.7% of P-224 output). In H-308 it is cooled from 59°C to the optimum hydrolysis temperature of 50°C. The Hydrolysis and Fermentation Schedule (Table 6.2.5.B) shows the sequence relationship between hydrolyzers and fermentors.

Cellulase enzyme produced in area 400 is added at the rate of 11,600 kg/hr. The slurry is agitated by two side-mounted agitators (A-307) providing 0.4 hp/1000 gal of stirring power. In addition, the slurry is re-circulated through a single bottom outlet and into three separate re-circulation lines. Each re-circulation line has a steam-warmed heater to maintain temperature of the slurry at 50 °C. Each line has an inlet 120 degrees of the others around the top of the tank.

The hydrolysis tank has a 30 degree cone bottom to ensure effective empting of the high solids slurry and has a volume of 1,419,529 L (375,000 gal). The cone bottom is supported by a full concrete foundation.

Each tank has a 3,000 gpm Discflo pump to accomplish the re-circulation of the high solids slurry. This pump turns the tank volume over once every two hours to avoid localized product inhibition and provide even temperature control. The slurry is divided up into three lines to increase diverse mixing once returning to the tank. The three warmer configuration was chosen because of concern for localized over-warming of the high solids slurry along the sides of the exchangers resulting in local denaturing of enzyme. There is no heat of reaction for hydrolysis[7] and it is possible that the heat capacity of the slurry and agitation power are sufficient to maintain the 50°C, however, this is unlikely hence the designed warming capacity described above. Table 6.2.4.B contains the calculations for enzymatic hydrolysis. All tanks and pumps are sterilized with hot caustic CIP between batches.

Table 6.2.4.B: Enzymatic Hydrolysis Calculations

41,484	gal/hr (cellulose and cellulase)	
24	hrs of stirring	
995,616	gal of stir cap. required	
375,000	gal/stir vessel	
3	number of vessels (add 1 for cleaning)	
90%	fill of vessels	
337,500	Operating volume	
9.0	time to fill (hr)	@ 691 GPM
2.0	empty time (hr)	@ 2,800 GPM

C. Cost Estimation

The hydrolysis area requires several pieces of equipment that were not included in the lignocellulosic model. However, the agitators from the lignocellulosic fermentors were used at full scale as the agitators for the hydrolysis tanks. The hydrolysis tanks (375,000 gal each) were scaled by using the new price quotes for the production fermentors (F-300) at 750,000 gal., scaled at 0.5 with a 2.0 installation factor to account for the cone bottom price difference and the extensive concrete foundation.

Discflo provided budgetary pricing of the 3,000-gpm re-circulation pumps. The hydrolysis warmers were priced based on recent Merrick experience as was the hydrolyzate cooler H-308.

6.2.5 Fermentation

A. Overview

Hydrolyzed stover slurry is pumped from the hydrolyzers, through coolers, and into the fermentation vessels. A recombinant *Zymomonas mobilis* developed at NREL performs co-fermentation of xylose and glucose. This co-fermentation does not (by process definition) include saccharification (SSCF). However, 20% of the original stover sent to hydrolysis remains unhydrolyzed after the 24 hrs. There is also cellulose present in the fermentation seed slurry which is added with as the inoculum. We have assumed (with confirmation from NREL[8], Dr. Jim Linden[6], and Dr. Ron Thomas[7]) that the cellulase still present in the hydrolyzate will provide SSCF with an estimated 39.5% conversion of cellulose to glucose over the 48hr fermentation time. This results in an overall conversion of cellulose to glucose of 84% as compared to 88% in the lignocellulosic model. This produces a slightly lower yield of ethanol per ton. However, the shorter combined hydrolysis and fermentation time of 72 hrs as opposed to 168 hrs translates to capital cost savings. Table 6.2.5.A compares the SHCF (900 metric ton/day) to the lignocellulosic model SSCF (2000 metric ton/day).

Table 6.2.5.A: Comparison of SHCF (900TPD) and SSCF (2000TPD*.45)

High Plains York Co-located Summary:	York Co-located	% of reference model	NREL Lignocellulosic "Reference Model"
DTPD (metric ton)	900	100%	900
stover (dry short ton/yr)	347,223	100%	347,223
ethanol (gal/yr) after rectification	25,746,124	97.7%	26,340,609
yield (gal/dry short ton)	74.1	97.7%	75.9
yield (gal/dry metric ton)	81.8	97.7%	83.6
hydrolysis + ferm. Time (hr)	72.0	42.9%	168
conversion of cellulose to glucose	84.0%	95.5%	88.0%
Additional EtOH (gal/yr)	(594,485)		

In addition to the glucose "wort" that is added to the fermentors, the *Z. mobilis* (fermentation seed) is also added along with corn steep liquor for nutrients, and ammonia for nutrients and pH control. The fermentation seed culture is initiated in the laboratory, as with the cellulase seed, and then transferred to multi vessel seed trains using detoxified stover slurry, ammonia, and corn steep liquor as the substrate.

The production fermentors are run batch-wise and at the end of each cycle are pumped to a beer well for surge control, then on to distillation.

B. Design Basis

Process Flow Diagram -P101-A301

Hydrolyzed stover is pumped from the hydrolyzers (T-307) by the re-circulation pumps through hydrolyzate coolers H-302 where the temperature is dropped from 50°C to 30°C. It then flows to the production fermentors, which are twice the size of the hydyrolyzers (see Table 6.2.5.B: Hydrolysis and Fermentation Schedule).

Process Flow Diagram -P101-A302

Simultaneous to hydrolyzate filling, the inoculum is added to account for 10% of the final fermentor working volume (see Table 6.2.5.C). This seed is grown in a series of five progressively larger vessels (F-301-5), each providing 10% inoculation to the next larger size. (see Table 6.2.5.D). Vessels 1-3 are jacketed, agitated package units and vessels 4 and 5 are agitated with cooling coils. As was mentioned with cellulase seed production; fermentation seed production residence time could likely be reduced by using hydrolysed slurry from the hydrolysers (T-307) which is already high in glucose, as opposed to using the unhydrolysed slurry directly from T-232.

The detoxified stover slurry is cooled with H-301 to 30°C in preparation for the seed production. For seed production, 24 hours in each vessel size has been determined by NREL to be sufficient for the cell count increase desired[11]. The seed is pumped to the seed hold-up tank (T-301) with pump P-302. The inoculum is pumped to the appropriate filling fermentor with P-301 as needed.

As with cellulase production, the number and configuration of seed vessels, which was chosen to be most appropriate, was that used in the lignocellulosic model with two trains of five vessels each (Table 6.2.5.E).

In the production fermentors, co-fermentation progresses for 48 hours while the hydrolyzed stover (sugar solution) is converted to an alcohol with a final content of ~5.3% (see Table 6.2.5.F for fermentation conversions as defined by NREL for lignocellulose feedstock[11]). It has been assumed here that this alcohol concentration is not high enough to significantly inhibit cell metabolism.

During fermentaion, cooling is provided by pump P-300 re-circulation through fermentation cooler H-300. The final beer is sent to the beer well (T-306) providing a constant flow to distillation with pump P-306.

All tanks and pumps are sterilized with hot caustic CIP solution between batches. Although the cellulase seed is of very low concentration with respect to the total volume of fermentation broth and *T. reesei* is a fungus - which tend to be slower reproducing than the ethanogenic bacteria – the *T. reesei* is added to fermentation in living form and hence represents an infection to the fermentation. The projected losses as a result of this infection need to be assessed for the separate hydrolysis and co-fermentation configuration. Physical, thermal, and chemical attempts to kill the fungus prior to use in hydrolysis are most likely detrimental to the enzyme and so therefore not attractive options. In our mass balance, 7% loss to infection is accounted for, leaving 93% of the sugars available for ethanol production.

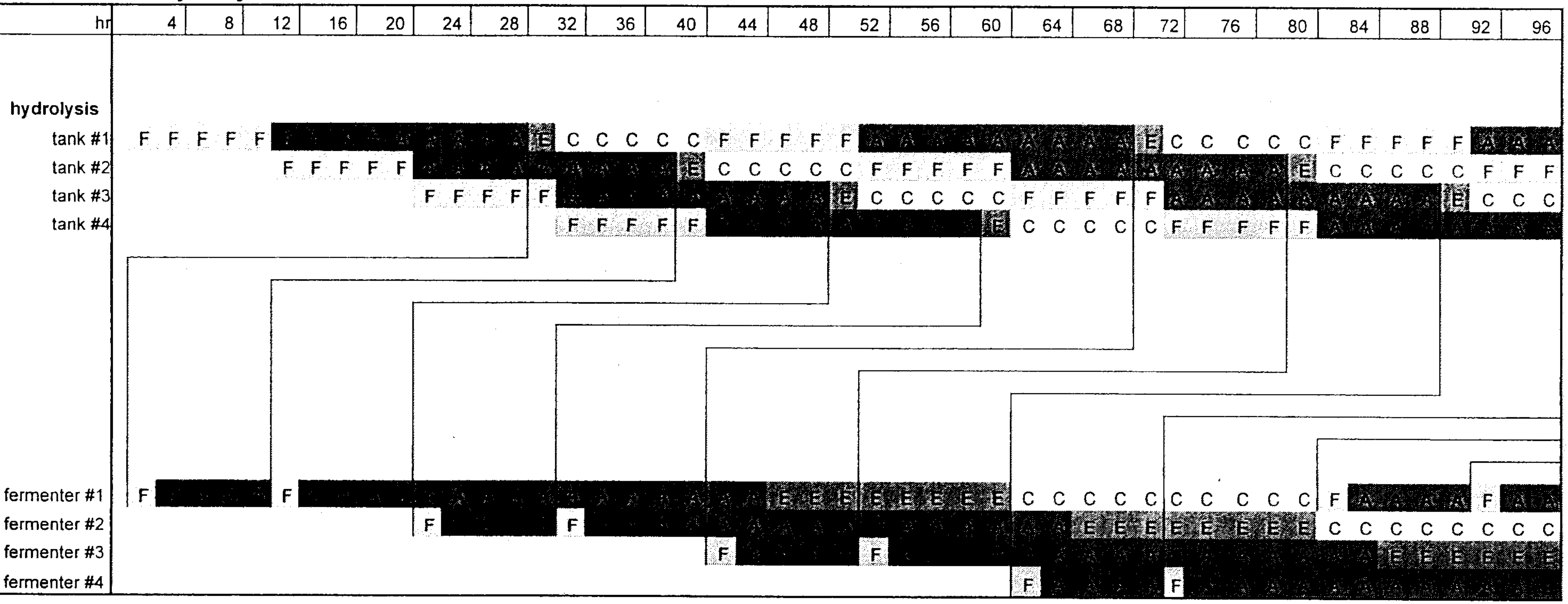
Table 6.2.5.C: Fermentation Conditions and Factors

fermentation		
time (hr)	48	
temperature (°C)	30	
hydrolyzate to fermentation (kg/hr)	157,136	89.7%
seed to fermentation (kg/hr)	17,529	10.0%
total flow to fermentation (kg/hr)	175,175	100.0%
% solids	8.1%	
CSL (kg/hr)	438	0.25%
Ammonia	71	0.04%

Table 6.2.5.D: Fermentation and Seed Production Design

	fermenters
46,246	fermentation broth (gal/hr)
2,219,812	fermentation volume (gal)
48	fermentation time (hr)
750,000	fermenter volume (gal)
90%	fill of vessels
675,000	operating volume
3.0	number of fermenters (add 1 for cleaning)
4.0	fill time/fermenter (hr) (hydrolyzate only)
771	empty pump rate to stripper (GPM)
14.6	empty time (hr)
	fermentation seed production
17,995	seed production broth flow in (kg/hr)
4,478	gal/hr broth in
24	batch time (hrs)
161,192	seed hold vessel (gal) (36hr)
67,544	inoculation to each fermenter (gal)
280	inoculation pump rate (GPM, for two hours out of every 10)
2	number of trains ("A" and "B")
80,596	vessel #5 operating vol. (gal)
89.6%	% working volume
90,000	vessel #5 capacity (gal)
9,000	vessel #4 capacity (gal)
900	vessel #3 capacity (gal)
90	vessel #2 capacity (gal)
9	vessel #1 capacity (gal)

Table 6.2.5.B: Hydrolysis and Fermentation Schedule

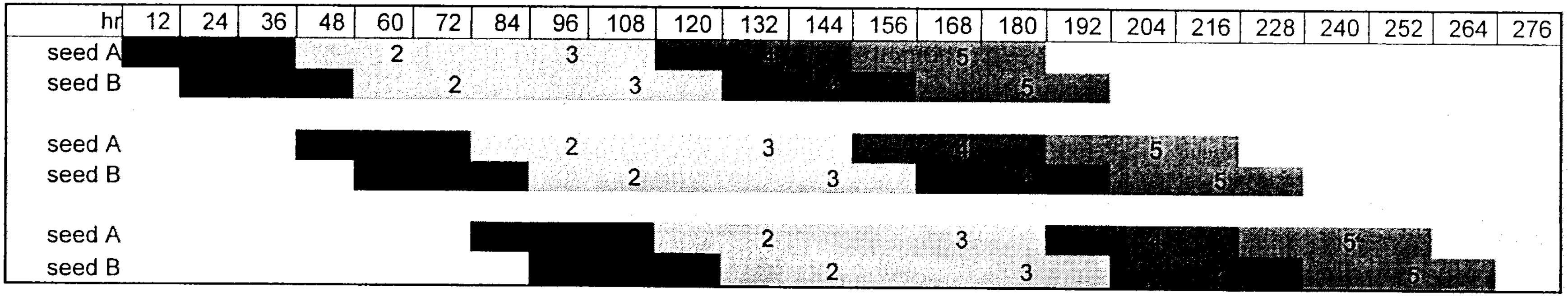


hydrolysis vessel size

hydrolyzer	375,000	gal
fermenter	750,000	gal

filling	F
active	A
empting	E
cleaning	C

Table 6.2.5.E: Fermentation Seed Production Schedule



vessel #5 capacity (gal)	90,000
vessel #4 capacity (gal)	9,000
vessel #3 capacity (gal)	900
vessel #2 capacity (gal)	90
vessel #1 capacity (gal)	9

Table 6.2.5.F: Fermentation Conversions

Glucose		→	ethanol	+ 2 CO_2	0.920
Glucose	+ 1.2 NH_3	→	6 *Z. mobilis*	+ 2.4 H_2O + 0.3 O_2	0.027
Glucose	+ 2 H_2O	→	2 glycerol	+ O_2	0.002
Glucose	+ 2 CO_2	→	2 succinic acid	+ O_2	0.008
Glucose		→	acetic acid		0.022
Glucose		→	lactic acid		0.013
ethanol + 2 CO2		→	ethanol		0.500
3 xylose		→	5 ethanol	+ 5 CO_2	0.750
xylose	+ NH_3	→	5 *Z. mobilis*	+ 2 H_2O + 0.25 O_2	0.029
3 xylose	+ 5 H_2O	→	5 glycerol	+ 2.5 O_2	0.002
xylose	+ H_2O	→	xylitol	+ 0.5 O_2	0.006
3 xylose	+ 5 CO_2	→	5 succinic acid	+ 2.5 O_2	0.009
2 xylose		→	5 acetic acid		0.024
3 xylose		→	5 lactic acid		0.114
% loss to contamination		→	lactic acid		0.070

C. Cost Estimation

The production fermentors received new budgetary quotes from Matrix Service, Inc. due to their high cost. Although the fermentation seed hold tank was over $100,000 it was only marginally so and is believed by Merrick engineers to be reasonable budgetary quote at $105,003. All other equipment in area 300 was cost scaled from the NREL lignocellulosic model.

6.2.6 Distillation and Dehydration

A. Overview

Separation of ethanol from the water/lignin slurry is accomplished via distillation (stripping and rectification) followed by dehydration to nearly 200 proof with molecular sieves. Gases coming off of the fermentors and fermentation seed vessels contain some ethanol in addition to various volatile organic compounds (VOCs). These gases are collected and sent to a scrubber where the VOCs are dissolved in cascading water with the non-condensable gases such as CO_2 being vented to the atmosphere. The water stream from the scrubber is pumped to the beer well for future distillation.

The existing York stripper (Beer Column) was originally designed for approximately twice the flow that it is currently handling. It was believed that this column could process the flow from both the grain plant and the stover plant without significant modification. However, the mixing of the recombinant fermented stover stream with the yeast fermented corn stream contaminates the still bottoms with the recombinant organism. The resulting distiller's grain looses its high co-product value due to market resistance against genetically modified organisms. Therefore, a stripping column is included in the equipment list. In addition, a "Kill Tank" has been added to maintain sterilization conditions long enough to ensure that the recombinant *Z. mobilis* is destroyed and not released to the environment.

The condensed vapor streams from the separate stripping columns could be combined for rectification. However, the rectification column and molecular sieve units at the York facility do not have the design capacity to process this quantity of feed. For this reason it was determined that the rectification and drying sections of the two processes would remain separate as well. The two streams combine at the existing High Plains alcohol Quality Assurance tanks.

B. Design Basis

Process Flow Diagram-P100-A501-3

Beer leaving the fermentation area is sent to the pretreatment area where flash vapors from T-203 preheat the beer in H-201 to 95°C. Once the heated beer travels to the distillation area it is heated once more in H-512 to 100°C using stripping column bottoms. The stripping column (D-501) separates the ethanol/water vapor from the lignin/water liquid.

The separation is accomplished with 32 actual trays, which are 48% efficient (NREL[11]). The feed is from tray 4 from the top. Trays are Nutter V-grid, which tolerate high solids with good efficiency. They are spaced 24 inches apart and the column is 7 foot, 6 inches in diameter. Overhead pressure for stripping and rectification is 26 psia. Overheads are removed to the scrubber and contain 100% of the CO_2 with 0.4% ethanol, 99% of which is recovered in the scrubber and recycled to the rectification column. A stream of 37% w/w ethanol to water is taken from actual tray 8 and fed to the rectification column. This represents 99% of the ethanol introduced to the column minus overheads mentioned and losses to the bottoms stream.

The stripping column bottoms are pumped using the beer column bottoms pump (P-501) to the kill tank (T-513). The temperature of 122°C with a designed 30 minutes residence time in the kill tank are sufficient to destroy the recombinant *Z. mobilis*. P-517 feeds the beer warming exchanger H-512 and pumps the sterilized stripping column bottoms to the evaporator system E-501-3.

The ethanol/water side draw from the stripping column is vapor fed to the rectification column (D-502). The rectification column separates the water and ethanol from this feed as well as the molecular sieve regeneration vapor and concentrates it nearly to its azetrope. This is accomplished with 60 Nutter V-Grid trays having 57% efficiency. The primary feed is on tray 44, with dehydration regeneration feed (higher in ethanol) on tray 19. The column is operated with 26 psia of overhead pressure and has a reflux ratio of 3.2:1. A 92.5% ethanol (w/w) stream is removed from the top of the column representing 99% of the ethanol that entered the column. The reflux condensing is provided by giving this energy up to drive the evaporators (E-501).

The overhead vapor from the rectification column is further "dried" using a Delta-T molecular sieve (M-503). These vapors are superheated and fed to the sieve columns where the water and a small amount of ethanol are absorbed. The sieve column is regenerated using a small slipstream of dried ethanol and a vacuum. This "wet" ethanol is sent back to the rectification column as mentioned earlier and the remaining pure, dry ethanol is sent to the existing High Plains York quality assurance hold tanks for testing with the other alcohol produced on site.

The CO_2 and off gasses from the fermentors and beer column are sent to the scrubber. The scrubber is a packed column using Jaeger Tri-Pack plastic packing, with 4 theoretical stages and well water to recover ethanol and other VOCs. This recycles ethanol and releases CO_2 with less than 40 short ton per year of organics. The water exiting contains 2.5% ethanol and is sent to the beer well for distillation with the beer.

All specifications for equipment in this area are from NREL[11].

C. Cost Estimation

Matirx Service, Inc. provided a budgetary quote for the kill tank. The kill tank bottoms pump (P-517) was scaled from the lignocellulosic beer column bottoms pump, which was designed for the same material. The distillation columns, although over $100,000, were cost scaled from the lignocellulosic model, as was the Delta-T molecular sieve. This was done because Merrick engineers believed that new vendor quotes on these complex, high cost items would not deviate significantly from those in the lignocellulosic model. All other equipment in this area was scaled from the lignocellulosic model.

6.2.7 Beer Column Bottoms Centrifuges and Evaporators

A. Overview

The stripping column bottoms are sent to a "kill tank" to assure that the *Z. mobilis* is destroyed with time at high temperature. The bacteria are not very heat tolerant and it may be possible that they are killed in the beer column and the kill tank may not be necessary, however, in this study it is included. The killed bottoms are then sent to a triple effect evaporation system where more water is driven off and the soluble and insoluble (lignin) solids are condensed. Energy to drive the evaporators comes from the rectification "heads" as mentioned in the distillation section above. The condensed solids are then centrifuged to remove remaining water and sold at fuel value.

The lignocellulosic model currently makes use of this lignin and syrup by burning it in a burner, boiler, turbogenerator. However, in this study it was our interest to consider the economics with respect to reduced capital with co-location and therefore eliminated this capital-intensive configuration. Depending on the revenue (or cost) value of the lignin as sold, this configuration may need to be reconsidered. This is in light of the fact that the stover facility steam requirement is too large to share existing boiler infrastructure at York and therefore a new boiler is needed anyway.

Water from the centrifuges is recycled at not more than 25% to avoid build-up of ions that produce osmotic conditions detrimental to the fermentation bacteria.

83.9% of the evaporator condensate is used as clean recycle water as compared to the existing York facility, which only recycles 25%. This difference is due to the design of the evaporators, which in the study case, use indirect contact of vapor and syrup which keeps the vapor clean and available for process water use.

B. Design Basis

Process Flow Diagram-P100-A504 and A601

Heat to drive the evaporators (E-501-3) comes from using E-501 as the rectification column reflux condenser. The lignin/water slurry is condensed in the first evaporator to nearly 11% insoluble and 3.5% soluble solids. It is then pumped using P-511 to the beer column bottoms centrifuge (S-601) in the wastewater treatment area (600).

All of the syrup (11.3% total solids) from the second and third evaporators is also sent to this area and sprayed on the centrifuge wet-cake (lignin solids at 34.7% total solids). Although this increases the water content of the wet-cake (now 26.5% total solids), this syrup cannot be recycled with the water stream if sent to the centrifuge because it is an important outlet for inorganic salts. These salts would otherwise build up to levels toxic to fermentation. Another option for removal of these salts, which needs to be considered in the future, is treatment as wastewater or the use of a burner/boiler configuration as with the NREL reference model.

Evaporated water from the evaporators is condensed in H-517 and pumped with P-514 to wastewater treatment (16.1%), ion exchange regeneration (48.5%), and acid dilution (35.4%).

Centrifuge wet-cake and syrup is conveyed via screw conveyor (C-601) to the lignin load-out bin M-614 which has a 15 minute capacity for rail car switching time. From here it is fed into rail cars and sold for its fuel value.

C. Cost Estimation

Although the beer column bottoms centrifuge has a cost much greater than $100,000 no new vender quote was obtained because it was felt by Merrick engineers that such action would not produce a significant change in cost. Therefore, the centrifuge, along with the evaporators and other equipment in this area was scaled and costed based on the equipment in the lignocellulosic reference model[11].

6.2.8 Area Requirements

The stover processing facility would likely be located on the north side of the existing corn facility (please refer to Appendix 2 "Interoffice Memo"). Presently this area is cornfield and is owned by the York facility. Feedstock handling, bale processing, and shredded stover storage would be located here requiring 7 acres (7,872 m^2). Pretreatment, fermentation seed production, hydrolysis and production fermentation would take place in a 4047m^2 building, which would be a mirror image of the existing fermentation building and located back-to-back with it to the north. Enzyme production would take place in an attached building with an area of 1,278 m^2.

Post ion exchange liquid would be pumped to the southeast corner of the existing facility for overliming near the distillers grain load out rail spur. The evaporators, lignin separation, and load out would be located here as well, having a 1,382m^2 footprint. This could make use of the existing rail spur for lime delivery and easy gypsum and lignin load out by truck or rail. However, due to the heavy load out traffic, an additional rail spur to the east of the overlimeing building would be more practical. The liquid would then be pumped back over to the mix tank (T-232) in the stover fermentation building.

The new distillation columns and mol sieves would be located between the existing fuel grade and industrial grade distillation areas. The additional cooling tower and chilled water packages would be located next to the existing north cooling tower. It is believed that the new boiler will fit into the existing boiler house.

Wastewater treatment would be placed along side the existing treatment area either to the north or the south.

An alternative to the above plan locates the entire stover facility to the southwest of the existing wastewater treatment area.

6.2.9 Utility and Chemical Requirements

6.2.9.1 Water

Process Flow Diagram-P100-A602 and A902-3

The plant water source is on-site well water. The estimate for the corn stover addition assumes that an additional 400 gpm well can be drilled. This provides sufficient make-up water for the facility, primarily required due to evaporation form the cooling tower. The facility is has zero water discharge with the exception of storm runoff water which is collected in a pond (M-108) and metered to waste water treatment. This water may be used as process water or discharged to the City of York wastewater treatment facility. For mass balance purposes, a high flow rate resulting from large storms was used to size the handling of this water. Therefore, the rate in streams 616 (storm pad run-off), 830 and 831 (flow thru wastewater treatment), and 617 (discharge to the city of York) will vary greatly depending on precipitation. Design basis flow assumptions and calculations can be found in Table 6.2.9.1. The detailed and summary water balance can be found at the end of Appendix 4.

Table 6.2.9.1: Storm Water Calculations

Bale receiving pad (ft^2)	250,000
precipitation (in/hr)	2
storm hours/wk	5.6
run-off to WW treatment (gal/hr)	311,688
storm run-off (gal/wk)	1,747,767
Flow through WW treatment (gal/hr)	10,403
kg/hr to WWT	39,407
Holding pond (one week)(ft^3)	233,643
dimensions 200 x 150 x 8ft (ft^3)	240,000

Wastewater was evaluated by Dr. Joseph Ruocco of Phoenix Bio-Systems, Inc. based on mass balance information provided by Merrick. His resulting report includes design explanation, configuration, mass balance, and operating costs as well as recommendations for further work. This report is included as Appendix 6.

Twice the cooling tower requirement of that existing at York is needed for the stover facility. Therefore a 40,000-gpm cooling tower system was included in the equipment list. This unit was scaled from the lignocellulosic model. 1,000 ton of additional chilled water capacity was also included. The York facility currently has 1 million gallon storage and pumping capacity for firewater. This was decided by Merrick engineers to be sufficient for the addition of the stover facility.

6.2.9.2 Ancillary Equipment

Process Flow Diagram-P100-A901 & A903

The clean in place (CIP) system from the lignocellulosic model as designed by Delta-T was included in the equipment list and scaled to 45%. All pumps, tanks and exchangers in areas 300, 307, and 400 as well as the evaporators, stripping column, kill tank, beer pre-heater and stripping reboiler are sterilized with hot caustic solution.

Plant air scale and pricing was used directly from the lignocellulosic reference model without scaling. However, it was decided by Merrick engineers that only one 500 cfm unit would be needed to augment the existing 1,000 cfm capacity.

6.2.9.3 Steam

Process Flow Diagram-P100-A801-3

The existing plant has two boilers. One is run near its capacity while the other is held at hot idle as an immediately available spare. However, the steam requirement for the stover facility is 174,187 lbs. per hour (Table 6.2.9.2). This is nearly twice the existing steam capacity at the York site, and so a new 200,000 lb/hr boiler was quoted by AALBORG Industries. This boiler produces 205 psig steam at 200°C (390°F) with 60,000 lb/hr going to 71°C (160°F) superheat.

Table 6.2.9.2: Boiler Calculations

methane energy BTU/ft^3	1,000
boiler eff.	0.85
incoming BTU/#	249.0
steam out BTU/#	1,199.6
delta H (BTU/#)	950.6
#/hr steam @ 389.9(F)	174,187
steam BTU/hr	165,589,610
superheat (F)	156.5
# superheat	40,077
superheat BTU	6,272,086
BTU consumed/hr	190,428,051
Ft3 methane/hr	190,428
methane #/ft3	0.04227
methane kg/hr	3,651

6.2.9.4 Fuel

The existing boilers are fueled with natural gas with the distillers grain driers fueled by natural gas and methane in the digestor off gas. The fuel cost for the co-located plant assumes that the stover addition will be fueled with natural gas as well and that methane produced will be sent to the driers as in the existing arrangement.

The lignin will be sold for its fuel value. Taking into account the Btu requirement for the heating and vaporization of the water (611 kcal/kg; 1,100 Btu/lb.) and a lignin energy content of 6,111 kcal/kg (11,000 Btu/lb), the gross annual fuel value is $7,848,926 at a rate of $2.50 per million Btu. This value is assumed to directly offset the cost of transportation to a customer - such as an electricity generation facility - where the lignin is used as boiler fuel. The capital estimate for this study does not include a solids fired boiler or steam driven turbine generator set as did the lignocellulosic reference model.

6.2.9.5 Power

Power for the currently existing plant and for the corn stover addition will be purchased from the local grid at a price of $0.35 per KW. Additional switchgear, substation, transformers, and motor control centers will be required and these have been included in the civil structural costs for the proforma. Power consumed by the stover plant was calculated as the sum of each user based on that users work duty. For equipment which was difficult to calculate work for (i.e. pretreatment reactor), a ~900 ton per day stover Aspen Plus model produced by NREL was consulted.

6.2.9.6 Chemicals

Process Flow Diagram-P100-A701

Chemicals required for the processing of corn stover include sulfuric acid, calcium hydroxide (lime), ammonia, corn steep liquor, antifoam (corn oil), sodium hydroxide (for clean in place – CIP), and gasoline (denaturant). In addition to these, a variety of cooling tower and boiler feed water chemicals currently in use at York were taken into account for use in the proforma.

The Pure Vision enzyme production technology requires a "small amount of low molecular weight enhancement factor." Due to the proprietary nature of this componant, it has been assumed that it will be delivered in drums or totes, handled with forklifts, and requires no special handling/storage precautions or procedures.

6.2.10 Transportation

Transportation of materials in and out of the facility is by two primary modes. These include road (truck or tractor) and rail. As mentioned in the Facility Description of this report, the York facility has several rail spurs adjacent to a Burlington Northern main line. Along the north side of the plant is US highway 34.

Corn stover will be received via highway speed rated tractors with trailers as described in section 6.2.1. Sulfuric acid will be received by rail car as will lime and corn steep liquor. Antifoam will be received by truck roughly every twenty-one days. Other chemicals (denaturant, ammonia, others) will be received by truck deliveries as is currently done at the facility.

Transport of products and waste from the plant will leave by rail. There is a significant amount of lignin solids to be sold (63,778 kg/hr) and this is loaded into rail cars from M-614. Transport of this material requires relatively continuous filling of eight rail cars each day (5650 ft^3 cars). This is very labor-intensive requiring three personnel per day.

Denatured fuel ethanol will also be shipped out via rail using the existing York facility infrastructure.

Gypsum waste from the overliming process is produced at the rate of 1,137 kg/hr. This requires the removal of 27,288 kg/day (60,158 lb/day) of waste. This is dropped into roll-off/roll-on dumpsters by the Hydrocyclone /Rotary Drum Filter (S-222). It is then shipped by truck to the county landfill. This requires three, four-axel trucks, every two days. The disposal cost associated with this is included in the proforma at $33 per short ton as quoted by High Plains Corp. Land application was discussed, however, we decided that the large quantity of gypsum produced annually would be detrimental to the land where applied.

Movement of material within the facility is with done with loaders, although forklifts may be used for intact bale transport. Forklifts existing at the York facility will be used for the transport of totes and drums.

6.2.11 Storage

Process Flow Diagram-P100-A701

Storage of materials at the corn stover facility requires considerable space, due the volumes of materials used. Stover is the greatest example of this, requiring 23,226 m^2 (250,000 ft^2) of concrete pad (M-102) for 4 days bale storage and handling. In addition to this there is 3 days worth of shredded stover storage (M-107) requiring 13,920 m^3 (600,000 ft^3) of volume.

A 240-hour supply of sulfuric acid is contained in a 75,758 L (20,000 gal) storage tank (T-703). There is an additional 7576 L (2000 gal) of sulfuric acid storage in the pretreatment area just prior to its use (T- 201).

Lime is stored in a 126m^3 (4455 ft^3) bin (T-220), which provides a maximum of 15 days storage and can be filled with 1.5 rail cars.

Corn steep liquor is stored in T-720, which is an 114,243 L (30,160 gal) vessel with 120 hours of storage. Antifoam is stored in a 45,455 L (12,000 gal) vessel (T-707) providing a 21 day supply. This large supply was chosen to take advantage of better economics by receiving via truckload (~9,000 gal) as opposed to multiple totes or drums.

Ammonia storage will make use of the existing tank at the York facility with the possibility of an additional tank as may be arranged by the plant manager and his vendor. Sodium hydroxide will be stored in the existing storage at the facility.

7 CAPITAL AND OPERATING COSTS

7.1 CAPITAL COSTS

The stover addition to the York plant can be constructed for approximately $79.4 million (including one month O&M operating costs for start-up) after an estimated 10% contribution of federal and state grants. The capital cost of the stover facility is strongly impacted by several important factors. These are in the areas of feedstock handling, pretreatment, various pumps and agitators, detoxification and wastewater treatment. Please see Appendix 5 for the equipment list. These areas each require high cost equipment, which with further research and systems development, could be significantly reduced. Civil engineering and other capital costs are outlined in Appendix 7. More detailed explanation and suggested ways of addressing these areas are outlined in section 12. Table 9.1 summarizes the financial assumptions.

Capital cost benefits of co-location with the High Plains York corn-to-ethanol facility are that there is no need to purchase land, and the road and rail accesses are pre-existing. In addition to this, the administration center and infrastructure are pre-existing. These facts help to offset some of the high cost equipment required.

Appendix 5 (the equipment list) has comparisons between the study equipment costs and the reference model as scaled to 45% (1999 costs with weighted average scaling exponents used). Also included is a comparison of the electrical workloads. The workloads and equipment costs are organized by area.

The comparison shows that the co-located study model equipment costs are $14.8 million less than the lignocellulosic model. This represents a 19.5% cost savings. There appears to be a $0.5 million capital savings by separating hydrolysis and co-fermentation. This is a result of fewer vessels required due to decreased residence time as noted in Table 6.2.5.A. The use of the PureVision cellulase production technology appears to result in a $1.7 million capital cost savings due to the reduced cellusase prduction scale required as a result of the higher specific activity of the enzyme.

Installation factors have been revised (in most cases increased) for the co-located study case and this has an effect on the installed costs (see Volume II of this report). A comparison of the installation factors and weighted average scaling exponents is also on the equipment list under each area.

7.2 OPERATING COSTS

Operating costs for the corn stover processing facility total $29.2 million a year, but with further development of the issues mentioned in section 12 "Recommendations for Further Work," these costs may be reduced. The costs are due largely due (aside from feedstock cost itself, which is by far the greatest cost center) to the system of feedstock handling and the labor that it requires. This accounts for $0.88 million/yr out of the total labor cost of $2.0 million/yr.

Electricity expenses are large ($3.8 million/yr). This is primarily due to energy needed for pumping and mixing slurries with assumed high viscosity, and aeration of wastewater treatment nitrification.

Chemical expenses add significantly to the operating cost with large quantities of lime ($0.83 million/yr), sulfuric acid ($0.72 million/yr), and ammonia ($0.60 million/yr). The result of the combination of these is that the facility has a negative cash flow of $185.3 million over its twenty-year life (assuming $35/dry short ton feedstock). More details of the operation assumptions can be found in Table 9.1, the proforma section 9, and Appendix 7.

The operation and maintenance costs are based on personnel required for the processing areas and the rates currently paid at the York facility. The management and overhead costs are modeled as a percentage of the operations labor costs. Some of the personnel, particularly in the maintenance and labor areas can be shared between the stover and corn facility. The operations experience of the corn facility personnel is one of the greatest benefits of the co-location configuration although this benefit is not included in the capital cost. Although some of the processing is different between corn and stover feedstocks, in general, the operational experience of the corn facility could result in very significant savings at time of plant start-up and the initial months of operation. Capital estimates do not account for start-up costs, which could include two to three months at less than full production plus unknown equipment retrofitting.

Chemical and energy rates per unit are those currently paid by the existing York facility. Wastewater quantity was determined by cutting in half, the volume of receiving pad runoff and charging at the rates used by the existing ethanol facility. The stover facility recycles all process water from wastewater treatment. Storm runoff water is sent to the City of York for treatment. There is no purchase of water due to the fact that the facility has on-site wells. Electricity consumption has been added to account for well operations.

The 4 principle operating costs, in order of greatest to less cost are:

1. Corn Stover ($12.2 million @ $35/DST)
2. Boiler Fuel ($4.0 million)
3. Electricity ($3.8 million)
4. Labor ($2.7 million including overhead)

8 TABLES of Important Design and Cost Factors

FEEDSTOCK HANDLING CALCULATIONS AND ASSUMPTIONS

bales (each)	
weight dry (#)	1200
weight wet (#)	1600
% solids	68.0%
length (in)	60
diameter (in)	70
% stalk (w/w)	82%
% cob (w/w)	18%

trucks (each)	
# / delivery (dry)	20,400
bales / delivery	17
delivery days	350

plant run time	
days / year	350
% of year	96%
hrs / day	24

feed rate		
kg / hr wet (101 PFD-A101)	71,977	
% solids	52.1%	
# / hr (dry)	82,662	37,489kg/hr (dry)
ton / hr (dry short ton)	41.3	37.5metric tons
ton / day (dry short ton)	992	900metric tons
bales / day	1,653	
ton / day (wet short ton)	1,323	1,200metric tons
ton / year (wet short ton)	347,182	
bales / year	578,637	
feedstock spec. size (in)	3/4 x 5/8 x 1/8	

deliveries		
bales / day	1,653	2,315design capacity (7/5)
bales / hr	138	193 "
trucks / day	97	136 "
weigh time / truck (min.)	10	10 "
delivery hrs / day	12.0	12.0 "
deliveries / scale / day	72	72 "
number of scales required	1.4	1.9 "

receiving and processing		
bale receiving pad (ft^2)	250,000	
dimensions (ft)	500	
forklift time per truck (17 bales) (min.)	30	
forklifts/loaders	5	6 "
bale processing	122	170 (wet short ton/hr)(12hr/day)

storage	
days of storage	3
shredded density (wet #/ft^3)	15
bunker volume (wet short tons)	4,376
bunker volume (ft^3)	583,499
(200x100x30) =	600,000

waste water run-off calcs.	
bale receiving pad (ft^2)	250,000
precipitation (in/hr)	2
storm hours/wk	5.6
run-off to WW treatment (gal/hr)	311,688
storm run-off (gal/wk)	1,747,767
flow through WW treatment (gal/hr)	10,403
kg/hr to WWT	39,407
holding pond (one week)(ft^3)	233,643
dimensions 200 x 150 x 8ft (ft^3)	240,000

loader fuel	
loaders	7
loader hrs / day	87.6
fuel usage (gal/hr)	8
gal/day	701
kg/day	1907

FEEDSTOCK COMPOSITION		
Design or Cost Factor	**Value**	**Unit**
Corn Stover Feed Rate	900	Bone dry metric tons / day
Cellulose	45.39	Weight %
Xylan	23.86	Weight %
Arabinan	2.00	Weight %
Mannan	0.00	Weight %
Galactan	1.11	Weight %
Acetate	2.11	Weight %
Lignin	18.53	Weight %
Ash	7.00	Weight %
Total	100.00	Weight %

PRETREATMENT REACTOR CONDITIONS	
acid concentration	0.5%
residence time	10 min.
temperature (°C)	190
solids in the reactor	22%

PRETREATMENT REACTOR CONVERSIONS		
Reaction	**Conversion**	
$(Cellulose)_n + n H_2O \rightarrow n$ Glucose	Cellulose	0.065
$(Cellulose)_n + m H_2O \rightarrow m$ Glucose Olig	Cellulose	0.007
$(Cellulose)_n + n H_2O \rightarrow ½ n$ Cellobiose	Cellulose	0.007
$(Xylan)_n + n H_2O \rightarrow n$ Xylose	Xylan	0.75
$(Xylan)_n + m H_2O \rightarrow m$ Xylose Olig	Xylan	0.05
$(Xylan)_n + \rightarrow n$ Furfural $+ 2n H_2O$	Xylan	0.10
$(Xylan)_n + n H_2O \rightarrow (Tar)n$	Xylan	0.05
$(Mannan)_n + n H_2O \rightarrow n$ Mannose	Mannan	0.75
$(Mannan)_n + m H_2O \rightarrow m$ Mannose Olig	Mannan	0.05
$(Mannan)_n + \rightarrow n$ HMF $+ 2n H_2O$	Mannan	0.15
$(Galactan)_n + n H_2O \rightarrow n$ Galactose	Galactan	0.75
$(Galactan)_n + m H_2O \rightarrow m$ Galactose Olig	Galactan	0.05
$(Galactan)_n + \rightarrow n$ HMF $+ 2n H_2O$	Galactan	0.15
$(Arabinan)_n + n H_2O \rightarrow n$ Arabinose	Arabinan	0.75
$(Arabinan)_n + m H_2O \rightarrow m$ Arabinose Olig	Arabinan	0.05
$(Arabinan)_n + \rightarrow n$ Furfural $+ 2n H_2O$	Arabinan	0.10
$(Arabinan)_n + n H_2O \rightarrow (Tar)n$	Arabinan	0.05
Acetate $\rightarrow$ Acetic Acid	Acetate	1.00
n Furfural $+ 3 n H_2O \rightarrow (Tar)n$	Furfural	1.00
n HMF $+ 3 n H_2O \rightarrow 1.2 (Tar)n$	HMF	1.00

Note: These reactions are modeled as occurring simultaneously. Therefore, products of one reaction, e.g., furfural, are not considered a reactant in another reaction. Degradation of xylan all the way to tar is accounted for as a reaction of xylan to tar. Degradation of furfural considers the furfural entering the reactor in the recycle water.

CELLULASE PRODUCTION SPECIFICATIONS AND CALCULATIONS			
yield (FPU/(g cellulose+xylose))	200		
productivity (FPU/(L*hr))	75		
specific activity (FPU/g protein)	800		
initial cellulose concentration	4%		
cellulase requirement (FPU/g cellulose)	15		
enzyme production broth (kg/hr)	13,384	33.7%	of reference model
enzyme production broth (gal/hr)	3,533		
production time / vessel (hr)	160		
size of production vessels (gal)	88,335	33.5%	of reference model
production vessel operating volume (gal)	70,668		
number of vessels in operation (add 3 for cleaning)	8		
% fill of vessel	80%		
time to fill vessel (hr)	20	12.5%	of production time

CELLULASE SEED PRODUCTION			
% inoculation of production vessels	5.0%		
volume of inoculant needed (gal/vessel)	3,533		
inoculant needed every	20.0	hrs	
batch time for each seed production (hrs)	40		
seed production vessel #1 (gal)	11	33.5%	of reference model
seed production vessel #2 (gal)	221	33.5%	"
seed production vessel #3 (gal)	4,417	33.5%	"
trains of vessels	3 - "A", "B", "C"		

ENZYMATIC HYDROLYSIS	
% insoluble solids (is)	15.0%
temperature (°C)	50
time per slurry (hr)	24
flow per slurry (kg/hr)	157,136
% conversion cellulose to glucose (hydrolysis)	80.0%
% conversion (SSCF 48hr)	39.5%
(overall hydrolysis conversion)	84.0%

HYDROLYSIS AND FERMENTATION CALCULATIONS

	hydrolyzers	
41,484	gal/hr	
24	hrs of stirring	
995,616	gal of stir cap. required	
375,000	gal/stir vessel	
2.9	number of vessels (add 1 for cleaning)	
90%	fill of vessels	
337,500	operating volume	
9.0	time to fill (hr)	691 GPM
2.0	empty time (hr)	2,800 GPM
	fermentors	
46,246	Fermentation broth (gal/hr)	
2,219,812	Fermentation volume (gal)	
48	Fermentation time (hr)	
750,000	fermentor volume (gal)	
90%	fill of vessels	
675,000	operating volume	
3.0	number of fermentors (add 1 for cleaning)	
4.0	fill time/fermentor (hr) (hydrolyzate only)	
771	empty pump rate to stripper (GPM)	
14.6	empty time (hr)	
183,467	beer well (gal.) (4hr res.)	
	fermentation seed production	
17,995	seed production broth flow in (kg/hr)	
4,478	gal/hr broth in	
24	batch time (hrs)	
161,192	seed hold vessel (gal) (36hr)	
67,544	Inoculation to each fermentor (gal)	
280	Inoculation pump rate (GPM, for two hours out of every 10)	
2	number of trains ("A" and "B")	
80,596	vessel #5 operating vol. (gal)	
89.6%	% working volume	
90,000	vessel #5 capacity (gal)	
9,000	vessel #4 capacity (gal)	
900	vessel #3 capacity (gal)	
90	vessel #2 capacity (gal)	
9	vessel #1 capacity (gal)	

FERMENTATION CONDITIONS

time (hr)	48	
temperature (°C)	30	
hydrolysate to fermentation (kg/hr)	157,136	89.7%
seed to fermentation (kg/hr)	17,529	10.0%
CSL (kg/hr)	438	0.25%
ammonia	71	0.04%
total flow to fermentation (kg/hr)	175,175	100.0%
% solids	8.1%	

FERMENTATION REACTIONS AND CONVERSIONS					
Glucose		→	ethanol	+ 2 CO_2	0.920
Glucose	+ 1.2 NH_3	→	6 *Z. mobilis*	+ 2.4 H_2O + 0.3 O_2	0.027
Glucose	+ 2 H_2O	→	2 glycerol	+ O_2	0.002
Glucose	+ 2 CO_2	→	2 succinic acid	+ O_2	0.008
Glucose		→	acetic acid		0.022
Glucose		→	lactic acid		0.013
ethanol + 2 CO2		→	ethanol		0.500
3 xylose		→	5 ethanol	+ 5 CO_2	0.750
xylose	+ NH_3	→	5 *Z. mobilis*	+ 2 H_2O + 0.25 O_2	0.029
3 xylose	+ 5 H_2O	→	5 glycerol	+ 2.5 O_2	0.002
xylose	+ H_2O	→	xylitol	+ 0.5 O_2	0.006
3 xylose	+ 5 CO_2	→	5 succinic acid	+ 2.5 O_2	0.009
2 xylose		→	5 acetic acid		0.024
3 xylose		→	5 lactic acid		0.114
% loss to contamination		→	lactic acid		0.070

KILL CONDITIONS		
Kill Temperature	122	Degrees Celsius
Kill Residence Time	30	Minutes

Comparison of SHCF (900TPD) and SSCF (2000TPD*.45)

High Plains York Co-located Summary:	York Co-located	% of reference model	NREL Lignocellulosic "Reference Model"
DTPD (metric ton)	900	100%	900
stover (dry short ton/yr)	347,223	100%	347,223
ethanol (gal/yr) after rectification	25,746,124	97.7%	26,340,609
yield (gal/dry short ton)	74.1	97.7%	75.9
yield (gal/dry metric ton)	81.8	97.7%	83.6
hydrolysis + ferm. Time (hr)	72.0	42.9%	168
conversion of cellulose to glucose	84.0%	95.5%	88.0%
Additional EtOH (gal/yr)	(594,485)		

FINANCIAL ASSUMPTIONS

Design or Cost Factor	Value	Unit
Reference Year	1999	
Plant Life	20	Years
On-stream Factor	0.959	%
Construction Period	1.5	Years
Startup Period	2	Months
Ethanol Selling Price (quoted by High Plains)	1.10	$ per Gallon
Owner Equity Financing	0.25	% of Fixed Capital Invest.
Loan Term	15	Years
Number of Annual Compounding Periods	1	1
Nominal Loan Rate Basis	7.5	%
Operator's Hourly Rate (quoted by High Plains)	16	$ / hr
Technician's Hourly Rate (quoted by High Plains)	16	$ / hr
Non-Skilled Laborers Hourly Rate (quoted by High Plains)	16	$ / hr
Supervisor's Hourly Rate (quoted by High Plains)	20	$ / hr
Payroll Overhead Factor	0.35	%
Operators / Day	14	
Technicians / Day	4	
Supervisors / Day	2	
Non-skilled Laborers / Day	7	
Purchased Electricity (quoted by High Plains)	0.035	$ / Kw*hr
Purchased Fuel Gas (quoted by High Plains)	2.50	$ / million BTU
Water	0	$ per thousand gallons
Water Disposal (quoted by High Plains)	1.00	$ per thousand gallons
Gypsum Waste Disposal (quoted by High Plains)	33.00	$ per short ton
Denaturant (quoted by High Plains)	0.375	$ / gallon

9 FINANCIAL PRO FORMA

A 25 million gallon per year corn stover-to-ethanol plant co-located at the High Plains York corn-to-ethanol facility does not appear to be an economically viable concept at this time. The stover addition to the York plant can be constructed for approximately $79.4 million after an estimated 10% contribution of federal and state grants.

The ethanol sale price was assumed to be $1.10 per gallon as quoted by High Plains Corp. This did not include the $0.10 per gallon federal small producers credit as the first 15 million gallons of annual production by the corn facility receives this credit and the stover facility addition will not fit the definition of a small producer at the combined production of 62.7 million gallons per year. The $0.54 per gallon federal excise tax credit to the blender was included for this base case. Table 9.1 summarizes the financial assumptions for evaluating this model. They are also discussed in section 7: Capital and Operating Costs. Appendix 7 contains the complete pro forma.

It was assumed in the base case that the lignin value just covered the expense of its transport to a purchasing facility. The purchaser is assumed to be an electricity producer. The value (and cost of transportation cost) of the lignin is estimated at $7.8 million per year. Discussion of this can be found in section 6.2.9.4.

With the feedstock cost of $38.59 per dry metric ton ($35 per dry ton) as quoted available in the York, NE area, the facility has a negative twenty-year net cash flow of $186.2 million dollars and a large negative internal rate of return (IRR). To get the IRR up to 1%, the stover cost was decreased to $15.93 per dry metric ton ($14.45 per dry short ton). **This scenario is considered to be the base case for pro forma and sensitivity analysis purposes.** Stover was adjusted because the only variables that will impact the economics greatly are capital cost, stover cost, and ethanol value. Merrick believes that a decrease in stover cost is more likely than an increase in ethanol value (i.e. to ~$1.38/gal achieving the 1% IRR). Adjusting stover cost is also more effective and more likely than a decrease in capital to improve the economics of this type of plant. In a facility which might purchase cellulase, the cost of this enzyme would be very significant and could be considered as well. This plant addresses this cost with on-site production. A discussion of the effect of on-site celllulase production using the PureVision technology, as opposed to the NREL reference model or purchased enzyme follows in section 11: Sensitivity Analysis "Comparison of Cellulase Sources".

The feedstock cost of $15.93 per dry metric ton ($14.45 per dry short ton) produces an IRR of 1% with a twenty year pre-tax net cash flow of $6.4 million dollars. Although this is positive cash flow with a positive IRR, in general the IRR needs to be closer to 20% to be considered economically attractive.

TABLE 9.1: FINANCIAL ASSUMPTIONS

Design or Cost Factor	Value	Unit
Reference Year	1999	
Plant Life	20	Years
On-stream Factor	0.959	%
Construction Period	1.5	Years
Startup Period	2	Months
Ethanol Selling Price (quoted by High Plains)	1.10	$ per Gallon
Owner Equity Financing	0.25	% of Fixed Capital Invest.
Loan Term	15	Years
Number of Annual Compounding Periods	1	1
Nominal Loan Rate Basis	7.5	%
Operator's Hourly Rate (quoted by High Plains)	16	$ / hr
Technician's Hourly Rate (quoted by High Plains)	16	$ / hr
Non-Skilled Laborers Hourly Rate (quoted by High Plains)	16	$ / hr
Supervisor's Hourly Rate (quoted by High Plains)	20	$ / hr
Payroll Overhead Factor	0.35	%
Operators / Day	14	
Technicians / Day	4	
Supervisors / Day	2	
Non-skilled Laborers / Day	7	
Purchased Electricity (quoted by High Plains)	0.035	$ / Kw*hr
Purchased Fuel Gas (quoted by High Plains)	2.50	$ / million BTU
Water	0	$ per thousand gallons
Water Disposal (quoted by High Plains)	1.00	$ per thousand gallons
Gypsum Waste Disposal	33.00	$ per short ton
Denaturant (quoted by High Plains)	0.375	$ / gallon

10 SENSITIVITY ANALYSES

The variables that can affect the overall economics of the co-located stover processing facility the greatest include ethanol sale price, stover price, capital cost, ethanol yield, cellulase cost, and lignin value. These factors were evaluated and graphs of the resulting IRR's were produced, with the exception of cellulase cost which was addressed separately with a comparison of options. No sensitivity for lignin value was run because it has been assumed that its value will just pay for its shipping to an end user. Should the co-located facility have a burner/boiler/turbogenerator for combustion of this lignin and production of electricity, it may become a fuel credit with additional electricity credit. This route was not taken in this project in interest of using a less expensive boiler and the low local cost of electricity. If this were to continue to be the assumption, a market for the lignin would need to be developed and could affect the overall economics positively.

The feedstock price used for the base case is \$15.93 per dry metric ton (\$14.45 per dry short ton) for reasons described in section 9: Financial Pro Froma. Feedstock price has a very significant impact on the economic feasibility of the stover processing facility. Ethanol value was assumed to be \$1.10 per gallon as quoted by High Plains Corp. Appendix 7 contains the complete pro forma.

The results of the sensitivity analysis show that if ethanol price were to get to \$1.20 per gallon (with the \$0.54 credit), the co-located facility could reach an IRR of nearly 11.5%. More realistic is the \$1.00 - \$1.10 per gallon range (and lower) at which point the facility is no longer profitable.

Feedstock price has the most impact on the economics of the facility as described in sections 9 and 10. The sensitivity analysis shows that if feedstock could be available for \$11.03 per dry metric ton (\$10.00 per dry short ton), the facility could reach an IRR of almost 7%. The feedstock price of \$38.59 per dry metric ton (\$35 per dry ton) determined by High Plains however, results in a very low economic feasibility. The facility does not have twenty-year positive net cash flow until feedstock cost go down to about \$16.69 per dry metric ton (\$15.14 per dry short ton).

Capital costs for the facility are very large with respect to the amount of ethanol produced at \$3.34 per annual gallon capacity (as a compared to a common \$2 or less for the corn to ethanol industry). It is believed by Merrick that the capital estimate could change a considerable amount if the issues outlined in Section 12: Recommendations For Further Work are resolved. Without these issues addressed, it is difficult to determine whether the change in capital will be an increase or decrease. Sensitivity analysis of capital investment shows that the IRR increases about 3% for each 10% decrease in capital cost (\$14.45 per dry short ton base case). Substantial grant funds in conjunction with cost savings of resolved issues could together improve the economic attractiveness of this co-located facility.

Perhaps of the greatest benefit to the facility would be to increase ethanol yield. The sensitivity analysis shows that if ethanol yield could reach 90 gal/dry short ton (from

74.15 gal/dry short ton), the facility could be economically attractive with an IRR of nearly 25% (assuming $14.45/dry short ton stover).

Comparison of Cellulase Sources
Of additional interest within the Building a Bridge-to-Corn Ethanol Industry, High Plains York, NE co-located corn stover-to-ethanol project, was the economic comparison of on-site cellulase production using PureVision technology to the reference case from NREL and the purchase of commercially available cellulase. This comparison illustrates the significant benefits of on-site production of cellulase, especially when the PureVision process is used.

The comparison was conducted by isolating the enzyme production processes for the reference case (scaled to 45% with SHCF) and the High Plains York tailored case (see Appendix 7, Cellulase Source Study, Comparison of On-site cellulase production methods, $per lb. calcs.). These processes were then each analyzed for mass and energy balance, equipment, utility requirements, raw and processing materials, financing, operations and maintenance, overhead, taxes, and depreciation factors to reflect the differences in the two scenarios. The resulting "cost of enzyme production" provides a reasonable approximation of the real cost associated with on-site cellulase production for each case.

It is important to note that the amount of cellulase required varies for the various cases not only because of the differences in specific activity (FPU), but also because each case has a different amount of cellulose to hydrolyze. For example, the purchased cellulase models require a 3.8% increase in the number of FPUs required at 15 FPU/g cellulose due to the fact that the cellulose that was used for cellulase production is now available for conversion to glucose then ethanol. It could be argued that this like "comparing apples and oranges" in that the amount of cellulose to be hydrolyzed is not consistent for all cases. However, this fact was accepted in this case because to make the amount of cellulose requiring hydrolysis equal in all cases would require a feedstock throughput change and therefore, entire facility resizing costing.

The result shows that in the NREL reference case, on-site production cost is $0.027/lb. of crude cellulase slurry. The cellulase produced with the PureVision process costs $0.022/lb of crude cellulase slurry. This equates to an annual cost savings to the facility of $1.2 million. A large portion of this savings is accounted for by the $1.7 million capital cost savings mentioned in section 7.1 in the form of lower property tax and debt retirement/depreciation. The remaining savings are principally accounted for by utilities, raw materials, processing materials, and O&M costs. There is also a decrease in ethanol production associated with this (1,179,071 L/yr [211,275 gal/yr]) because the cellulose that is sent to produce the larger volume of cellulase required is no longer available for conversion to glucose and then to ethanol. Appendix 7 contains the details and a summary comparison of the cellulase production options. Any processing or equipment cost changes that may result from this decrease in production are not taken into account in the economic comparison, because they are thought to be relatively small. However, the impact on revenue is accounted for. For economic and equipment descriptions of this scenario, please see Appendix 7, Cellulase Source Study, Comparison of On-site cellulase production methods.

The International Filter Paper Unit (FPU) is a commonly accepted way to measure the specific activity of an enzyme. Specific activity is the rate at which the enzyme converts a substrate to a product – in this case, cellulose to glucose. The NREL reference model enzyme has a specific activity of 600 FPU/gram of protein whereas the PureVision cellulase has a specific activity of 800 FPU/gram of protein. The result here is that 25% less enzyme is required for the same degree of cellulose conversion when using the PureVision cellulase production technology. Therefore, if cost of production per FPU is considered, the cost for the reference model enzyme is $4.60 per million FPU (MFPU) whereas the cost of the PureVision enzyme is $3.32/MFPU. The result in this application is a savings of $1.2 million per year in capital, O&M, utilities, overhead and other costs.

The above two on-site cellulase production models were then compared to the scenario with purchased, commercially available enzyme. The High Plains tailored case was altered to eliminate cellulase production equipment and included adjustments to all the other variables associated and mentioned in the previous analysis. Provisions were accounted for in terms of cellulase receiving and handling. It should be noted that the purchased enzyme is received in a concentrated, purified form and has shipping charges of $3.00 per mile for an average estimated distance of 750 miles (with tanker trucks hauling 50,000 lbs this calculates to $0.413 per lb. of cellulase). Thus, purchased enzyme has a much higher cost at $2.41 per pound (delivery cost included). Appendix 7, Cellulase Source Study, Comparison of On-site and Purchased Cellulase, contains details of two methods used to do this comparison (page 19) as well as and equipment list and summary sheet for each method.

It is perhaps more appropriate to base its comparison to the previous two cases on its FPU content than on a dollars per pound basis. Even this may not be the best basis for comparison until tests with the purchased enzyme can be run to determine its true activity (FPU rating) on the corn stover substrate. In addition, there is debate as to whether the FPU is even an accurate measurement unit for enzyme specific activity. However, the FPU basis will be used here providing the most accurate comparison considering the immature state of activity assay data on pretreated stover for all three enzyme sources.

Many factors go into trying to make an accurate comparison between these cellulases such as the substrate they are produced on, and the proportions of exo-nuclease, endo-nuclease, and especially β-glucosidase. These factors are likely the principle technical arguments supporting the uses of cellulase produced on-site using the same substrate as that which is to be hydrolysed.

Two methods for projecting the cost of purchased cellulase were used with vastly different results, but with identical conclusions. It should be noted that either method is only a best estimate for the reasons mentioned above and to follow. The first method "BASED ON PUREVISION LABORATORY RESULTS OF COMPARISON" is based on laboratory results obtained by researchers for PureVision[5]. Out of several purchased enzymes compared to the PureVision produced cellulase, the Specialty Enzymes Inc. Liquicell 2500 was used for our performance comparison basis. The second method "BASED ON PRODUCT SPECIFICATIONS PROVIDED BY SPECIALTY ENZYMES INC." is based on product specifications for Liquicell 2500 provided by Specialty Enzymes Inc.[14]. Specialty Enzymes Inc. also provided a purchased bulk cellulase price quote of $2.00 per pound which was used for both projection methods. Transport cost was assumed, as stated above, by Merrick.

The amount of enzyme (on an FPU basis) required for conversion of the cellulose is higher (3.8%) for purchased cellulase cases because the cellulose lost to the growth and production of *T. reessei* and cellulase is now available for conversion to glucose and then ethanol. This results in an increase in ethanol production of 933,825 gallons annually. Both purchased cellulase case calculation methods take into account the increase in revenue which results from the additional production. The equipment list for the purchased cellulase case is the same for both methods of projection and has been modified form the base study case as mentioned earlier. The section of Appendix 7, Cellulase Source Study, Comparision of On-site and Purchased Cellulase, Method A, "Equip. (purchased)" shows the equipment differences. There has been no adjustment to capital and operating cost to adjust for the additional ethanol production (i.e. larger fermentors or more Z. mobilis seed) because only 3.8% of the pretreated slurry was used for cellulase production. Merrick believes that this will be an insignificant additional cost.

METHOD A: "Based on PureVision Laboratory Results of Comparison"
Data based on tests performed by PureVision using their enzyme and several commercially available cellulases on hydrolysis of high-grade waste paper and low-grade restaurant waste paper have been performed[7]. The results show that the PureVision enzyme is 6.43 times more effective on the high-grade waste paper in an 18 hr period (see Appendix 7, Cellulase Source Study, Comparision of On-site and Purchased Cellulase, page 19 for this calculation). This could also thought of as having a specific activity of 125 FPU/g protein (as compared to 800 FPU/g protein for the PureVision cellulase), although a comparison on the FPU/g protein basis was not determined through tests. Pretreated corn stover has different characteristics (such as higher lignin content) and therefore it is likely that the multiple calculated is inaccurate when treating corn stover. However, being the only available results, they will be used for our purposes here.

The results of our comparison using this purchased enzyme assessment method show that the facility will require the delivery of 325,810 truck loads per year to supply the required amount of enzyme. The additional cost to the facility over the base study case (PureVision on-site cellulase production) is $4,484,964,258 annually. Clearly this is in no way a feasible option, both in terms of logistics and economics. However, it does illustrate the importance of on-site cellulase production. Appendix 7, Cellulase Source Study, Comparision of On-site and Purchased Cellulase, pages 14-29 contain details of the calculation which led to these conclusions and a side-by-side comparison of the cellulase source evaluation.

METHOD B: "Based on Product Specifications Provided By Specialty Enzymes Inc."
Product specifications were provide by Specialty Enzymes Inc. for a currently available cellulase used in the textile industry Liquicell 2500. The specifications for this cellulase include a higher specific activity than that which was found in the PureVision comparison

tests mentioned above, but was also not necessarily intended for use in hydrolysis of waste paper.

Using this method to project the effect of on-site cellulase production vs. purchased cellulase is much less dramatic than Method A, but is just as impractical, both logistically and economically. It shows that the facility would require the delivery of 21,637 truck loads of cellulase annually. This also results in an additional cost of $489,256,883 annually over on-site production using the PureVision cellulase production technology. The section in Appendix 7, "Cellulase Source Study", "Comarison of On-site and Purchased", pages 24-29 contain the details of this projection. The equipment list is the same as that for Method A.

SUMMARY

The comparision of PureVision and NREL reference model on-site enzyme production to the purchased enzyme illustrates the importance of efficient, high specific activity enzyme production. Achieving the high specific activity needed seems to be intimately related to an "acclimation factor" which comes from producing the cellulase using the same substrate as the cellulase is to hydrolyse (i.e. to hydrolyse corn stover efficiently, produce the cellulase on a corn stover substrate). It is likely that this acclimation factor is related to the relative proportions of exo-nuclease, endo-nuclease, and especially β-glucosidase[6,7,13]. The purchased cellulase has a cost of between $2,753.93 and $182.89 per million FPU, depending on the source of the specific activity specification (i.e. laboratory test or vendor). This is as opposed to the $4.60 and $2.32 per million FPU for the refenence model and base cases respectively. As mentioned before, the "FPU" unit used in these purchased specifications may well not perform the same as an "FPU" as defined in the reference and base cases (although this seems to be in contradiction to the supposed definition of an FPU).

Most important to note is that in making the comparison to purchased enzymes, a standard basis for comparison must be established. This is difficult due to the variety of substrate conditions, characteristics of the enzymes, and preparation forms. It is highly possible that the results of the comparison with purchased cellulase are not accurate due to the fact that no real comparison can be drawn between the purchased enzyme and crude on-site produced enzyme without a direct head-to-head comparison of both enzyme sources on actual pretreated corn stover. Such comparison should include identical assay and reporting protocols, more accurate and universal measuring unit definition (i.e. FPU/g protein), and identical cellulase forms (i.e. liquid).

11 CONCLUSIONS

The purpose of this report was to explore the business potential of producing fuel ethanol from corn stover at a facility co-located with an existing corn-to-ethanol facility. In doing so, a process design was selected and a mass balance was produced. From this, capital and operating costs were determined for co-location at the High Plains York ethanol production facility. An economic comparison between various cellulase enzyme sources was also evaluated. Important considerations were availability of low cost feedstock, sizing of the stover facility, on-site production of enzyme, and separation of hydrolysis and co-fermentation. Merrick also compiled a Pro Forma for the co-located plant, identified parameters that most significantly impact production costs, and performed sensitivity analyses on those parameters.

As an outcome of this effort, a conclusion has been drawn that while requiring a feedstock cost much lower than that available in order to remain profitable, there are enough areas for continued development that once addressed, may make this co-location a feasible option for corn-to-ethanol facilities.

A very important consideration should be given to the source of enzyme for a biomass to ethanol facility of this general design. The cost of purchasing available cellulases (which are not intended for biomass digestion) is extremely costly. On-site production of cellulase is a requirement, not an option, for an efficient facility of this size and design. The PureVision cellulase production technology represented here has a critical positive effect on the economics of this facility as compared to the NREL reference model and almost certainly to the purchase of currently available commercial cellulase enzymes.

Important technologies needing further development are outlined in section 11 "Recommendations for Further Work." In general these include feedstock handling, pretreatment, detoxification, enzyme production, and co-product value. Also worth closer consideration is the separation of saccharification and co-fermentation.

The fact that the co-fermentation is achieved by use of a genetically modified organism should be of great consideration for co-location opportunities. If the fermentative organism were not genetically manipulated, the two fuel ethanol production streams could possibly combine before product quality assurance, perhaps sharing stripping, rectification and/or dehydration resources (assuming the lignin could be marketed with the distillers grain). However, the use of genetically modified organisms may have a very negative impact on the marketability of the grain if combined. Public concerns and permitting with respect to the local presence of a facility relying on recombinants may not be positive as well.

The stover processing facility adds an additional 70% product to the York plant production. With an increase of this magnitude and the high steam and chemical requirements of delignification/hemicellulose hydrolysis, the over-scale of the existing York infrastructure would have to be very significant to avoid adding all new equipment (i.e. boiler, cooling tower, chilled water, wastewater treatment, rectification, and dehydration capacity). Therefore, in this case, little existing equipment can be shared by the two facilities.

Infrastructure, that can be shared include management, some personnel, operations experience, and some plot space as well as road and rail-sidings.

Sensitivity analysis shows that stover cost at $38.59 per metric ton ($35 per short ton) renders the facility uneconomic. This is the price quoted as available in the York, NE area by High Plains Corp. If stover were available at $16.69 per metric ton ($15.13 per short ton), the plant would "break even." Therefore, the facility does not appear to be economically attractive at this time. However, if the issues outlined in section 11 were reconciled, the economics of the co-located stover-to-ethanol facility would improve considerably.

In conclusion, this study shows that a corn stover-to-ethanol facility co-located with the existing High Plains Corp., York, NE corn-to-ethanol facility could be economically attractive (IRR=20%) if stover were available for $0.62 per metric ton ($0.56 per dry short ton). This is not a likely occurrence. Other variable changes, such as an increase in ethanol selling price, or establishing a gypsum value, are very unlikely considerations for increasing economic viability. The variable that needs to be focused on is the decrease in capital and establishment of a market for the lignin and to a lesser extent CO_2. On-site enzyme production using the PureVision enzyme production technology is one example of a good way to reduce the capital cost of a biomass-to-ethanol facility.

12 RECOMMENDATIONS FOR FURTHER WORK

The following recommendations are intended to direct further research into the operations and technology that we believe will have a great bearing on the economic feasibility of co-located lignocellulosic fuel ethanol facilities utilizing corn stover as feedstock. The ideas presented are only suggestions and have not been researched or studied in any depth as a portion of this project.

12.1 Feedstock Handling

The corn stover collection system, which requires stover to be baled, may not be the most efficient. It is very capital and labor intensive. If corn stover is not baled then the manual debaling costs would be eliminated and the expense of plastic netting and its disposal would be avoided. However, storage of stover in a loose form could be problematic.

Combining of the washing of dirt from bales (if this is extensively required) with storm water runoff could create a significant wastewater handling issue as well as a loss of feedstock. Our assumption here has been that the bales will require minimal washing via pressure hose spraying of visible contaminants on bales before they are manually debaled. Experience at NREL with their Process Development Unit indicates that there is bacterial contamination of pretreated biomass slurry, indicating that the pretreatment process is not sufficient for sterilization of soiled feedstock[8]. Therefore, soil entrained within the bales (i.e. root systems collected durring windrow raking and baling from the ground) is a potential source of infection and therefore a loss of production. In addition, the granular nature of the soil, if not removed, is detrimental to the structural integrity of critical equipment such as the pressure screw feed mechanism on the pretreatment reactor. A feedstock washing process was not thoroughly investigated for this project because we felt that a revised harvesting and feedstock handling system could reach a more practical and efficient solution, to the above-mentioned obstacles.

Harvesting with some size reduction directly from the combines as they harvest the grain should be considered. The stover could then be transported clean - without dirt and with minimal field debris - in bulk compressed loads to the facility. With additional size reduction, it could then be compressed in large silage type bins[10]. The feedstock could then be conveyed directly to a day bin via pneumatic conveyor and from here to the pretreatment reactor. Regional collection centers with bunkers may also play a role in this design.

Another consideration is to produce rectangular bales of the common dimensions 1.52m x 1.52m x 2.44 meters. These do not require the plastic wrapping that the round bales require and may make bale handling less tedious. Other possibilities include the pelletization of the stover at regional collection centers and storage at the ethanol facility in silos.

12.2 Pretreatment

The Pretreatment reactor is a very significant portion of the capital cost considering the fact that it is considered a single piece of equipment. The continuous flow configuration also has serious safety hazards associated with the high pressure feeding method. Other pretreatment system should be considered, for example a batch reactor system scheduled to provide an apparently continuous flow may be appropriate.

12.3 Detoxification

Ion exchange and overliming may not be necessary and their elimination or alternatives should be pursued. Perhaps this could be accomplished with alkali pretreatment prior to the pretreatment hydrolyzer for removal of acetate. Another consideration may be the use of a microbe with greater tolerance to the "toxic by-products" of lignocellulosic pretreatment such as a yeast (as opposed to a bacterium). This could save millions of capital and operating dollars in wastewater treating[9], purchased chemicals, and waste production (gypsum). Please refer to Appendix 5.

If ion exchange cannot be eliminated, it may be better to use sodium hydroxide rather than ammonia or ammonium hydroxide for ion exchange resin regeneration. Nitrification of ammonia in the waste water system is very expensive and may justify the use of the more expensive sodium hydroxide reagent.

12.4 Slurry Properties

There is a difficult balance between high solids content and low ethanol concentrations in the beer. The quantity of steam sent to the stripping column reboiler to heat the high volume of low ethanol concentration beer (5.3% w/w), represents 42% of the total low-pressure steam usage. This translates to large capital cost for stripping column, reboiler, natural gas consumption and a larger boiler. However, in order to increase the alcohol % of the beer, minimal water should be present in the liquor. This translates to high solids concentrations with difficult pumping and mixing characteristics.

This report assumes the successful use of centrifugal pumps in the pre-hydrolysis and hydrolysis sections of the plant to move high insoluble solids concentration streams. These pumps should be tested with actual flowing materials to guarantee that they are capable of the flow rates and discharge heads required by the process. Similarly, various vessels in the process contain mixtures having very high insoluble solids content. These vessels are assumed to have effective agitation to prevent settling, maintain temperature or maintain reaction. The service conditions are not common for conventional agitators and their effectiveness is yet to be proven. The use of helix pumps may be a point of investigation for this issue.

12.5 Enzyme Production and Use

The development of the hydrolysis enzyme is in its early stages. At this time, research has shown the technology to be effective for use on waste paper. We strongly suggest that additional work needs to be done in proving the requirements for enzyme growth and its efficiency in hydrolyzing corn stover. Areas of importance with regard to enzyme are the continued effort to increase specific activity and the establishment of an accurate industrial standard and assay method by which to compare all biomass-degrading cellulases such as the FPU or an amended version thereof.

12.6 Separate Hydrolysis and Co-fermentation

The timesavings, and therefore capital savings, that are suggested as a result of this study and especially the separation of hydrolysis and co-fermentation need to be verified on a larger scale under industrial conditions to confirm this benefit. Use of a non-genetically modified organism should be pursued due to lack of acceptance of recombinant organisms by the public, and associated management/containment issues.

12.7 Replacing Existing Capacity

Replacing a portion of the existing facility capacity with stover feedstock, as opposed to adding the capacity on, may save capital costs. The pretreatment areas for the stover and corn would need to be different, but the two streams could be combined after saccharification and the commonly used yeast could be used for C_6 fermentation. This would by-pass the five carbon sugars, however the issues associated with using the recombinant organism would not exist. All equipment down stream of saccharification could be used by both feedstocks. Another alternative would not combine the streams, but uses the same equipment for separate stover and corn fermentations. Variations on these and the impacts on co-products such as DDG need to be studied.

12.8 Use of Stillage

It may be possible to use whole or thin stillage for nutrients in enzyme production and fermentations instead of corn steep liquor. This needs to be studied as does the general nutritional requirements of the enzyme production and ethanol production fermentations.

12.9 Production of Grain Neutral Spirits

The production of a higher-grade industrial ethanol should be investigated. The higher value of the neutral spirits could significantly improve overall stover-to-ethanol plant economics. Product quality standards for this are very high, but may justify the extra expense of meeting them.

13 REFERENCES

1. Q.A. Nguyen et al., "NREL/DOE Ethanol Pilot Plant: Current Status and Capabilities," Bioresource Technology, 58(189-196) 1997.

2. "Modified Process Model Results for a Feedstock Composition Reflecting Corn Stover", National Renewable Energy Laboratory, Technical Memorandum, April 26, 1999.

3. Glassner, D. A.; Hettenhaus, J. R.; Schechinger T. M., "Corn Stover Collection Project," in *BioEnergy '98- Expanding Bioenergy Partnerships: Proceedings, Volume 2*, Madison, WI, pp. 1100-1110, 1998.

4. Phillippidis, G. P., "Cellulase Production Technology" in *Enzymatic Conversion of Biomass for Fuels Production, ACS Symposium Series 566*, Chapter 9,.

5. Ed Lehrburger, Pure Vision Technologies, Inc., Personal Conversations, February, 1998 to November, 1999.

6. Dr. Jim Linden, Colorado State University, Personal Conversations, April, 1999 to November, 1999.

7. Dr. Ron Thomas, Clemson University, Personal Conversations, April, 1999 to November, 1999.

8. Kelly Ibsen, National Renewable Energy Laboratory (NREL), Personal Conversations, May, 1999 to December, 1999.

9. Dr. Joseph Ruocco, Phoenix Bio-systems, Inc., Personal Conversations, October, 1999 to November, 1999.

10. Thomas Schechinger, Iron Horse Custom Farming LLC, Personal Conversation, July 26, 1999.

11. Wooley, R.; Ruth, M.; Sheehan, J.; Majdeski, H. and Galvez, Adrian, "Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis Current and Futuristic Scenarios," National Renewable Energy Laboratory, Technical Report, Golden, CO, July, 1999.

12. J. McMillan, "Bioethanol Production: Status and Prospects", *Renewable Energy*, 10(295) 1997.

13. Spindler, Dianne; Torget, Robert; Grohmann, Karel, "Ethanol Production by Simultaneous Saccharification and Fermentation (SSF) of Pretreated Corn Cobs and Corn Stover," Final Report for FY89 – Task #BF983231.

14. Dr. Vic Rathi, Specialty Enzymes and Biochemicals, Personal Conversation, December 21, 1999.

15. Ranatunga, T.D.; Hervis, J.; Helm, R.F.; McMillan, J.D.; Hatzis, C., "Identification of Inhibitory Components Toxic Toward Zymomonas mobilis CP4(pZB5) Xylose Fermentation," Applied Biochemistry and Biotechnology, January 23, 1997.

16. Claussen, E., "Acid Recovery and Recycle by Solvent Extraction," Office of Fuels Development 1996 & 1997 Project Summaries, Biofuels Program, http://www.esd.ornl.gov/bfdp/doeofd/96_97summaries/bioprogram.html

APPENDICES

1. **Feedstock Description Report, Task A1**
2. **Trip Reports**
3. **Cellulase Enzyme Dosage Study**
4. **Process Flow Diagrams**
5. **Equipment List**
6. **Waste Water Report**
7. **Proforma and Sensitivity Analysis**

Appendix 1

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February 1, 2000

TO: Fran Ferraro, Merrick & Company

CC: Greg Heuer, Chris Standlee

Report, Task A1, Feedstock Description

Project No. 19013442 Building a Bridge to the Corn Ethanol Industry

This report summarizes the results of research conducted to 1) determine the availability of corn stover, and 2) evaluate the spent distiller's grains (DDG), for conversion to ethanol at High Plains Corporation's York, Nebraska Ethanol Facility. References are cited where appropriate.

CORN STOVER

From consultation and literature available, the best economic area of collection was assumed to be within close proximity to the plant operations. For practical application, including primarily ease-of-access to major highways (Highways 81 and 34 and Interstate 80), this report covers a five-county area centered around York County. The Ethanol Facility is located on Highway 34 approximately 3 miles east of the interchange with Highway 81 and 7 miles north of Interstate 80. This area includes irregular boundaries, but will represent an approximately 70-mile maximum transportation route from field to collection warehouse to plant site.

In all cases, the most conservative data or estimates were used. The following table summarizes the tons of stover that could reasonably be collected, stored, processed and transported to the York facility. The 1997 – 1998 *Nebraska Agricultural Statistics* report[1] on "Corn For Grain" acres harvested for the crop years 1995, 1996, and 1997, revealed that 1995 resulted in the lowest acres (and yield). The University of

Nebraska has reported[2] on collectible corn residue for 25 counties including the 5 counties of interest in this report. Their data included low, high, and "best estimates", and provided for exclusion of Soil Conservation Acres. This report used the lowest reported data less the tons of Soil Conservation residue.

High Plains Corporation (HIPC) has received privileged information indicating that 60% removal of stover from fields is both economically and practically viable using a proprietary system of custom harvesting, baling, storage and transportation[3]. Assuming that 50% of those producers with stover available will contract to participate in a collection process, then 30% of the collectible stover would be available for conversion. It has been variously reported that up to 3.7 tons per acre of stover is available[3]. Ranges of reporting could result from the inclusion or exclusion or the cobs and shucks with the stalks (Iowa State University has reported[4] that cobs and shucks make up 1.0 tons per acre). The table also indicates the resulting tons of stover if 30% of the available corn-growing acreage participated and 2.0 tons per acre can be harvested (a randomly selected, conservative number that approximates a value provided by the proprietary custom harvester noted above). This comparison provides a range that may be used when evaluating conversion options and equipment requirements for the facility.

County	Corn Acres Harvested 1995 Crop Year Bushels	Collectible Stover 1993 Residue Tons	30% Acreage Participation @ 2.0 Tons/Acre	30% of Collectible Stover, Tons
York	242,000	249,000	145,200	104,700
Hamilton	250,400	305,000	150,240	91,500
Seward	125,100	86,000	75,060	25,800
Fillmore	180,800	186,000	108,480	55,800
Polk	150,800	137,000	90,480	41,100

Total Tons for Biomass Conversion	**569,460**	**to**	**318,900**

STOVER COLLECTION AND COSTS

Proprietary data[3] provided to High Plains Corporation indicates that this volume of stover can be harvested, baled, and transported to collection centers within 120 days of harvest at a delivered price of less than $35.00 per ton. Initial foray into this new feedstock at this volume will likely prove more costly until the collection centers and infrastructure are established.

DGS

The York facility uses approximately 13,800,000 bushels annually of grain to produce 36,000,000 gallons of anhydrous ethanol. The Distiller's Grains and Solubles (DGS) by-product will contain both insoluble portions of the spent grain combined with a portion of the soluble portions. The total plant output of soluble and insoluble solids (dry matter basis) is approximately 350 tons per day (124,250 annual tons). Testing analysis[5] indicates that 9% of this product is fermentable (enzyme soluble carbohydrate) and another 9% is fiber[6] that may be converted using cellulase technology. This equates to 63 tons per day of fermentable feedstock. Conversion of this 18% portion of the DGS to ethanol would also raise the protein level, which may add value to the remaining by-product. Conversely, addition of unconverted starches from the stover process along with the residual lignin and ash to the DDGS will significantly reduce the protein value.

CONVERSION – PLANT SIZE EVALUATION

Proponents[3] of various conversion technologies have professed to achieve up to 80% conversion of cellulose and hemi-cellulose to glucose, which equates to 135 gallons of ethanol per ton of biomass. Others have stated 75 gallons of ethanol per ton as a realistic goal. NREL has reported[7] that Corn Stover is 41% cellulose and 21% hemi-cellulose. If the conversion technology results in comparable corn/milo conversion, and the known corn/milo yield is 80 gallons of ethanol per dry ton of grain with 68% fermentable starches (2.6 gallons per bushel) then a ratio can be established to calculate theoretical stover yield. This relationship is shown in the following table.

For project evaluation, it is recommended that the conservative figure (or the average of the two assumptions) be used.

62 % "fermentables" in corn stover
68 % fermentables in corn/milo (starch, DMB)

80 gallons/ dry ton of corn/milo yield

64 gallon/ton assuming corn stover conversion is 80% of starch

$$\frac{62}{68} = \frac{X}{80}$$

$X = 73$ gallons/ton assuming conversion is equal

68 Gallons/ ton average of assumptions

Using this yield and the stover available from this research then 21,685,200 gallons can be produced from stover and 1,520,820 gallons from DDGS of anhydrous alcohol production.

References Cited

[1] *NEBRASKA AGRICULTURAL STATISTICS*, 1997 – 1998, Nebraska Dept. of Agriculture, Issued November 1998

[2] *Feasibility of Corn Residue Collection in Kearney, Nebraska Area*, Report of Findings for Western Regional Biomass Energy Program by University of Nebraska-Lincoln, Industrial Agricultural Products Center, September 1993.

[3] Summary notes of proprietary discussions with confidential business associate(s) by HIPC Plant Manager, D. Allison, 1998 - 1999.

[4] Report on Agricultural Residue Harvest and Collection, prepared for Western Regional Biomass Energy Program Western Area Power Administration by North Dakota State University, November 1994.

[5] Enzyme Development Corporation analysis report to Greg Heuer, HIPC Director of Plant Operations, October 19, 1998.

[6] Servi-Tech Laboratories Feed Analysis Report to HIPC, typical.

[7] NREL Memo, "Results of analytical analysis on fiber samples" to Danny Allison, HIPC from Kelly Ibsen, NREL, July 21 1998.

Appendix 2

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TRIP REPORT

DATE:	**April 1 and 2, 1999**
PROJECT:	**Building a Bridge to the Corn Ethanol Industry**
PROJ. NO.:	**19013442**
LOCATION:	**Iron Horse Custom Farming, Harlan, Iowa** **High Plains Corp. Ethanol Plant - York, Nebraska**

ATTENDEES:	**Danny Allison**	**High Plains Corp.**
	Joe Casey	**High Plains Corp.**
	Dale Bender	**High Plains Corp.**
	Tom Schechinger	**Iron Horse Custom Farming, LLC**
	Dick Voiles	**Merrick & Company**

Visit to **Iron Horse Custom Farming in Harlan, Iowa**:

1. The corn stover harvest, last fall, was cancelled by Great Lakes Chemical at just under 50% completion. Great Lakes could not sell their products.
2. Because the stover demand has fallen from a forecast of 65,000 tons/year to about 30,000 tons/year, Iron Horse is selling some of their equipment. A high speed tractor in good conditions is valued at approximately $60,000 to $65,000..
3. Great Lakes produces furfural, furfural alcohol, and Furfafill (a by-product used as a glue extender in fiber board) at their Omaha plant.
4. Great Lakes burns about 50% of the Furfafill by-product for energy. For each 20 tons of stover a ton of ash is produced. This ash was originally sent to landfill but now most is applied to local fields.
5. Iron Horse was successful in changing Iowa law to allow custom hauling using the high speed tractors. These tractors have air ride, air brakes and other safety features. They are more stable and much easier to control than conventional tractors. These tractors are superior to trucks in collecting corn stover because they are better able to work in snowy, wet or heavily frosted fields. They are more economical than trucks within about 40 miles of the delivery point. The advantage for farm tractor/trailers averages $1.49/ton of stover within the 40 mile haul distance.
6. Many factors influence the collection of corn stover. Farms near river bottoms would like to remove essentially all of the stover. However, conventional methods allow a pick up of 60 to 70 % before the amount of dirt inadvertently picked up becomes excessive. Fields on hill sides generally yield less stover and leave much of it on the

field to prevent erosion. Approximately 40% of the stover must be collected to make the operation economically attractive.

7. Collected corn stover is put into bales by multiple, independent baling contractors. Although there are numerous baling contractors, experience has shown that only about 30% of these are reliable and have the skill to make good, dense bales that will transport economically and store well.
8. If bales are not dense the transportation costs become uneconomical. Large round bales should be about 1200 pounds dry weight. Skill in making the bales can vary this weight by as much as 600 pounds.
9. Large round bales wrapped in plastic netting for transportation and storage have advantages over twine wrapped bales. Bales held by either sisal or plastic twine do not store well and allow losses from the bale at highway speeds.
10. Setting up a collection program is time consuming. Farmers need to understand the benefits for their individual farms and be convinced to try stover collection. Each set back (such as cancelled harvest) undermines the trust that must be established. There are often several benefits for a farmer besides the actual price paid for the stover such as being able to get into the field earlier in the spring, saving on disking operations, offsetting some increase in fertilizer cost by savings in the soil nitrogen addition requirements.
11. Additional benefits will happen once the program is shown to be successful. For example, there are hybrids that produce the same corn yield but have more foliage – leading to more stover. These may become attractive to farmers who don't want them now.
12. To do a harvest effectively in the short time window available means that one must be over-equipped. Practical use of the equipment will require the harvesting of other materials not having the same schedule. This should be part of the overall plan. For example, switch grass could be harvested after the corn stover harvest is complete.
13. Farmers would be more comfortable if they had more than a single buyer for their product.

Visit to the Harlan terminal of Great Lakes Chemicals:

1. The terminal had approximately 40,000 bales stored on 8 to 10 acres surrounding a processing building. Of the 40,000 bales, 22,000 were from the most recent harvest. Bales are stored in rows, stacked three bales high. Dense bales, wrapped in plastic netting were storing well. Some of the bales were from the previous year's harvest (that is, they have been in storage since the fall of 1997). Low density bales and bales wrapped in twine were falling apart and could not be moved as a bale.
2. The processing center chops the raw stover and extrudes it into about 1-1/2" diameter by 4 to 6 inch long pellets or bricks. These are conveyed into large trucks for transport to the Great Lakes' Omaha plant. They have experimented with the extruder and found that they can vary the density of the pellets to meet plant requirements.

3. Examples of the several products which could be manufactured solely or partly from corn stover were available. Included were fiberboard, animal bedding, seeding mulch, furfural etc.

York Ethanol Plant:

1. Toured the plant with the potential addition of a corn stover/spent distillers grain (SDG) addition foremost in mind.
2. The beer column can handle nearly twice the current load thus potentially eliminating the need to duplicate for new throughput.
3. A single boiler can easily handle the average steam requirement. However both boilers are run continuously in a turned-down mode in case one should fail. If a new plant did not add a third boiler this standby or spare capability would be lost.
4. The air compressor may be adequate to handle a second plant.
5. Chilled water systems may be adequate for a second plant.
6. When able High Plains sells wet spent distillers grain cake to local feed lots thus saving the cost of drying. If SDG must be transported a significant distance, it is necessary to dry it.
7. Dry SDG can be loaded into rail cars using a horizontal auger that evenly loads the car. Trucks are loaded with a front loader.
8. A set of P&IDs, block flow diagram and descriptive information were given to Merrick for project use.
9. The high quality ethanol distillation section was shut down due to lack of product demand. This situation is not likely to continue.
10. There appears to be ample space for plant additions either as a separate plant or as an integrated plant. Feed stock storage must be separately evaluated.
11. High Plains has targeted the week of April 5, 1999 to supply a draft report to the project for their parts of the corn stover project. Some of the most important aspects are:
    - Assumed available corn stover is 30% of the produced corn stover in York County and the two adjacent counties. This is roughly equivalent to a 70 mile radius. It means that 400,000 to 500,000 tons per year of stover is available.
    - Placing a value on delivered stover is not easily done. One approach would be to back into the highest cost stover could be for the operation to be economically attractive.
    - Based on 77 gallons of ethanol production from each ton of corn stover, the plant capacity will be roughly 36 million gallons per year. Published ethanol yields from corn stover vary from 75 to 135 gallons/ton. The value of 77 was selected to match an NREL paper that equated corn stover yield to 62% of corn grain yield.
12. NREL (Kelly Ibsen) has a consistent set of utility prices for the York plant, which she is using in developing an ASPEN Plus™ model. These utility prices should be used for this project.
13. York recently added a RO unit on boiler feed water, in which, will reduce the cost for this commodity. RO is due to high silica in the feed water.

14. Plant raw water is from three wells located on-site.
15. SDG is produced at the rate of 300 tons/day.
16. Some important questions which we must address during the course of this project are:
    - SDG is sold for $100 per dry ton as animal feed. Is this not too high a value for ethanol feed?
    - If SDG is fed to the ethanol plant presumably the volume of solids would be reduced as cellulase breaks down much of the fiber. However, there may be no effect on the proteins which are the basis of the live stock feed price. It may be that the value per ton as animal feed will increase if SDG is processed without stover?
    - There is a belief that the corn plant and the stover plant cannot merge until after distillation. Is there a basis for this?
    - The 300 tons per day of SDG yields 2 million gallons of ethanol per year. Does this small yield justify the cost of investigation in this study?
    - An alternative to increasing the overall ethanol production at York is to blend stover into the existing plant feedstock by backing out corn grain and observing the economic effects.

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TRIP REPORT

DATE:	November 3, 1999
PROJECT:	Building a Bridge to the Corn Ethanol Industry
PROJ. NO.:	19013442
LOCATION:	High Plains Corp. Ethanol Plant - York, Nebraska

ATTENDEES:	Brian Pasbrig	High Plains Corp.
	James Atmore	High Plains Corp.
	Dale Bender	High Plains Corp.
	Dick Voiles	Merrick & Company

On November 3rd, 1999 I traveled to York, Nebraska to visit the High Plains grain to ethanol production facility. The purpose of this visit was to discuss the potential placement of equipment in a new corn stover facility that would be built and operated at the same location.

The following is a compilation of notes taken during the visit.

Met Dale Bender (operations manager). Mr. Bender set up a meeting with Brian Pasbrig and James Atmore to discuss the various questions with me.

I expalined that my initial layout located the new facilities North of the administration offices in the cropland owned by High Plains.

- The bale receiving area (500 ft. x 500 ft.) would be located adjacent to the N/S road to the east. – The bale receiving area should have a separate access entrance with separate weigh in/out scales due to the current truck traffic entering the plant. Discussed having a separate access road from highway 34 for the bale receiving area and believe that this is possible, however the county would be less likely to approve an exit from highway 34 due to the number of trucks anticipated.

- The stover feedstock storage area was discussed. –The loader for this area should be included in the capital cost estimate as sharing a loader with the existing facility would be impractical due to the usage.

- Discussed the hydrolysis/fermentation building location and layout. – Location N. of the current fermentation building looks good. The building layout should be mirror image to the existing building with respect to tanks. Areas should be designated for control room, QC Lab, operator lab, and offices. Integration with the existing DCS system will need to be incorporated into the project, and possibly a central control room for both plants will need to be installed. The existing ammonia tank can be shared. The sulfuric acid tank should be added for a new facility. These tanks are presently loaded by truck.. The loading facilities can be used and transferred to the new facilities. The fermentor seed tanks should be located near the fermentors. The celulase enzyme production systems should be located in a separate building to the north.

- The material handling systems were discussed along with a new rail spur for lignin loadout and lime handling.

  - The current DDG rail spur (running N/S) might be extended north to allow use for lignin loadout, however the amount of rail traffic anticipated and rail car staging would likely interfere with the current truck traffic for grain unloading. This option could be studied further but at present does not appear to be feasable. A new rail spur to the ease of the DDG Loadout may be more practical.

  - Locating the evaporator and centrifuge area near the existing E/W rail spurs and pumping the slurry across the existing facility was also discussed. The lignin handling area could then be located east of the existing DDG facility and could use the DDG loadout spur (or supplement the spur) with minimum impact on the existing truck traffic.

  - The existing centrifuge building has spare locations for additional centrifuges. Locating the lignin centrifuges in this building would save significant capital costs due to the building costs being eliminated.

  - The lignin area should have a surge pit for conveyor upsets.

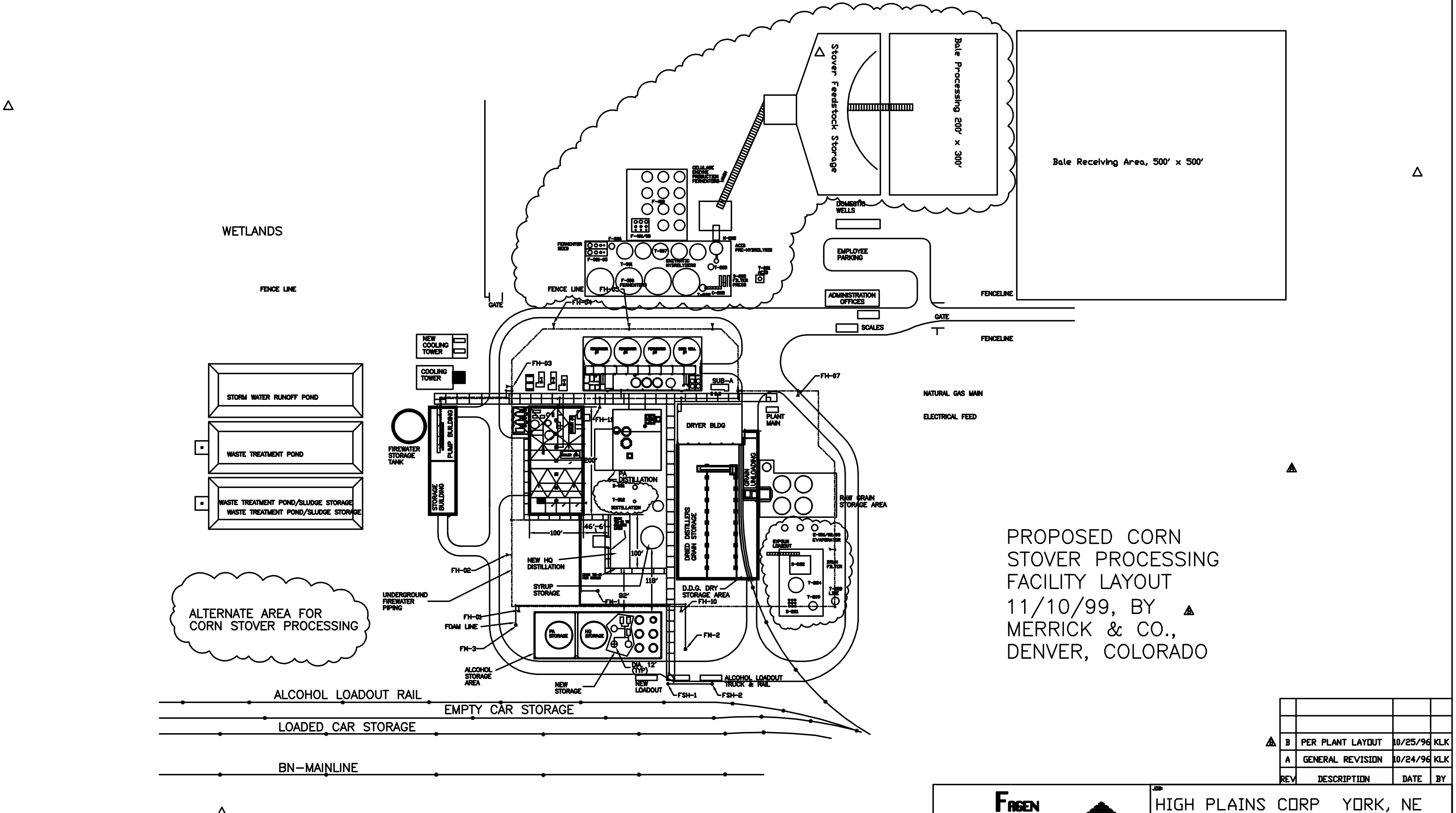
- The existing distillation area and mol sieve could be expanded to allow processing of the new stream. In particular if a preliminary separation was made (to say 160 proof) then the existing facility could probably handle the final refining. The movement of slurry across the plant becomes more attractive if the existing distillation facilities can be expanded to handle the extra capacity.

- The gypsum and lime handling would then be located near the lignin loadout area.

- Plant walkdown also revealed an area used for equipment laydown that could be used. An alternate layout will be produced that shows the new facility in the SW corner of the plane with access from the west (road one mile west). This option would allow minimal impact to the existing operation.

- The existing electrical capacity of the plant was discussed with Mike Kriewald. The line capacity to the substation should be adequate for the new facility, however a new 34.5 KV to 480 V transformer would be required for the new plant.

Appendix 3



REV	DESCRIPTION	DATE	BY
B	PER PLANT LAYOUT	10/25/96	KLK
A	GENERAL REVISION	10/24/96	KLK

FAGEN ENGINEERING LLC

Civil - Structural
Mechanical - Electrical
Engineers

(320) 564-4573
501 W. Hwy. 212, P.O. Box 159
Granite Falls, MN. 56241

JOB: HIGH PLAINS CORP YORK, NE

LOCATION: YORK, NE	DRAWN BY: D.C.	CHECKED BY: D.H.
ENGINEER: FAGEN ENGINEERING, LLC	SCALE: 1" = 50'	DATE: 9/5/96
CONTRACTOR:	JOB NUMBER	SHEET NUMBER
DESCRIPTION: SITE PLAN	EG1013	21-01

Cellulase Enzyme Dosage Study
Jim Linden
28. July 1999

I have reviewed literature published during the past ten years that describes the effects of cellulase enzyme dosage on extend and time dependence of hydrolysis of pretreated lignocellulose. The data has been collected with the purpose providing a recommendation regarding over-capacity of enzyme production for the Separate Hydrolysis and Fermentation (SHF) process under consideration. Ten relevant papers were found; the important facts from each will be reviewed in order of chronological appearance.

Comparisons of enzyme dosage and *Trichoderma* enzyme manufacturer with the hydrolysis of pretreated aspen wood was presented by Schwald et al. in 2 to 60 L reaction vessels (1). Over 85% of the cellulose could be hydrolyzed to glucose in 96 hours when an 8% substrate concentration was used with 9 FPU/g substrate. The same average conversion appeared to be complete in 48 hours when 17 FPU/g substrate was used. A visual estimation can be made from the attached figure (schwfig1).

Two related papers by Spindler, Wyman and Grohmann from NREL appeared in 1990 that described Simultaneous Saccharification and Fermentation (SSF) of dilute sulfuric acid pretreated herbaceous crops, which included corn stover (2, 3). Little difference was found in final yield between the low and high cellulase enzyme loadings. In all cases, SSFs showed faster and higher conversion than SHFs for the same enzyme loadings. For instance, comparison of 13 and 26 IU/g cellulose loadings with beta-glucosidase supplementation, corn stover SSF theoretical yields after seven days were 86 and 92 percent, respectively. Table 1 from reference 2 is duplicated as an attachment (spintab1).

A 1994 paper by Penner and Liaw provided some kinetic modeling for the *Trichoderma* cellulase system (4). Under conditions of substrate inhibition using high ratios of substrate to enzyme, the relative enzyme hydrolysis rate varied only 30 micromol/h over the range from 10 to 100 FPU / g microcrystalline cellulose substrate.

A paper in 1996 contained exactly the kind of information desired (5). The effect of enzyme loading on ethanol production in batch SSF of pretreated sugar cane bagasse using 7, 15 and 30 FPU / g cellulose was given in Figure 2, which is attached (lynnfig2). Ethanol production plateaus after 50 hours using 30 FPU / g. Treatment with 15 FPU / g had produced approximately 80% that of former case in 50 hours and 100% that of the former case in 300 hours. These data indicated an advantage of using the greater loading in SSF. Presumably similar results would be obtained in SHF. However, when examining the conversion based on cellulose concentration, the decline in final substrate utilization with declining enzyme loading was small. The effect was thought possibly to be pretreatment dependent, rather than a substrate-specific effect that might result from reduced inactivation of enzyme owing to the low lignin content of the pretreated material.

An alkaline pretreatment of corn stover was studied in a 1998 paper by Belkacemi et al. (6). Saccharification of the pretreated material was followed by fermentation, hence SHF. Indeed 55-70% of the cellulose was hydrolyzed after 48 hours, and the extent of hydrolysis was dependent not only on cellulase units, but also more dramatically on the amount of beta-glucosidase added to the system. This finding supports data of Spindler et al. (2, 3) that is presented above. Increasing the solids loading to 10% (w/v) during hydrolysis from 5% almost reduced the saccharification by half.

Baker et al. from NREL continued studies on enzyme mixtures using purified enzymes in a 1998 paper (7). Results revealed that at least one synthetic mixture utilizing enzymes from three different organisms delivers performance competitive with that of a "native" *T. reesei* system.

In conclusion, increasing the enzyme dosage by a factor of two appears to reduce the time to similar extent of conversion by from 10% (2) to50% (1) to 75% (5). The range encompasses different substrates and different enzyme systems. Certainly, using an enzyme system with sufficient beta-glucosidase reduces the advantage. Also, using easily convertable substrates, such as corn stover, reduces the advantage. Knowing that the cost of enzyme production contributes very significantly to the product value, I would find it prudent to use 15 FPU/g cellulose for SHF, especially since the enzyme produced on pretreated corn stover should have superior characteristics for hydrolysis of the same substrate (8).

Files: I:\process\3442\report\dosage.doc,spintab1.doc,schwfig1.doc

Bibliography

1. Schwald, W. et al. 1989. Assessment of Pretreatment Conditions to Obtain Fast Complete Hydrolysis on High Substrate Concentrations. Appl. Biochem. Biotechnol. 20/21:29-44.

2. Spindler, D. et al. 1990. Ethanol Production by Simultaneous Saccharification and Fermentation (SSF) of Pretreated Corn Cobs and Corn Stover. Proceedings of Corn Utilization Conference III, St. Louis, MO, June 19-21.

3. Spindler, D. et al. 1990. Evaluation of Pretreated Herbaceous Crops for the Simultaneous Saccharification and Fermentation Process. Appl. Biochem. Biotechnol. 24/25:275-286.

4. Penner, M.H. and Liaw, E.-T. 1994. Kinetic Consequences of High ratios of Substrate to Enzyme Saccharification Systems based on *Trichoderma* Cellulase. In Enzymatic Conversion of Biomass for Fuels Production, (M.E. Himmel, J.O. Baker and R.P. Overend, eds). ACS Symposium Series 566, American Chemical Society, Washington D.C., pp. 363-371.

5. Van Walsum, G.P. et al. 1996. Conversion of Lignocellulosics Pretreated with Liquid Hot Water to Ethanol. Appl. Biochem. Biotechnol. 57/58:157-170.

6. Belkacemi, K. et al. 1998. Ethanol Production from AFEX-Treated Forages and Agricultural Residues. Appl. Biochem. Biotechnol. 70/72:441-462.

7. Baker, J.O. et al. 1998. Hydrolysis of Cellulose Using Ternary Mixtures of Purified Cellulases. Appl. Biochem. Biotechnol. 70/72:395-403.

8. Philippidis, G.F. 1994. Cellulase Production Technology: Evaluation of Current Status. In Enzymatic Conversion of Biomass for Fuels Production, (M.E. Himmel, J.O. Baker and R.P. Overend, eds). ACS Symposium Series 566, American Chemical Society, Washington D.C., pp. 188-217.

Files: I:\process\3442\report\dosage.doc,spintab1.doc,schwfig1.doc

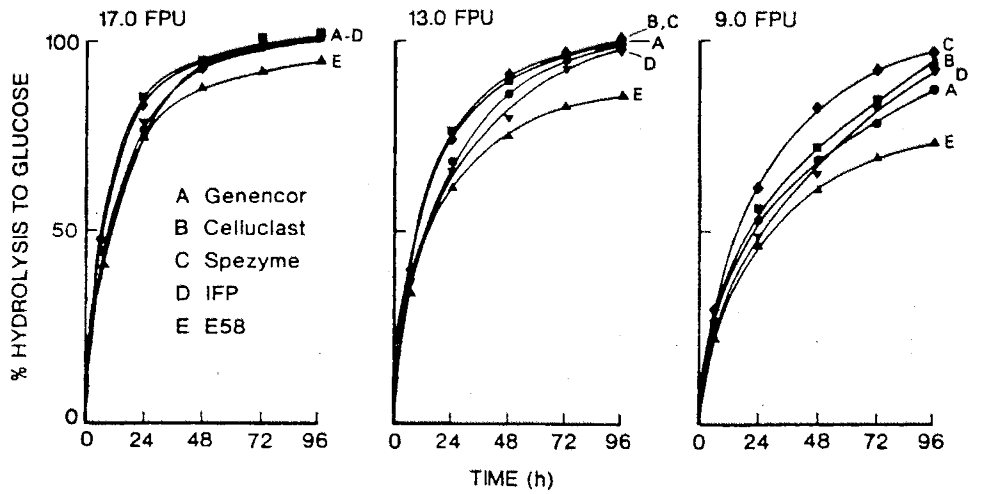


Fig. 1. Effect of enzyme concentration (FPU/g substrate) on enzymatic hydrolysis of pretreated aspen wood using various enzyme preparations supplemented with β-glucosidase (Novozym) to a constant level of cellobiase activity.

Tab. 1. SSF - Final (7 day) percent theoretical yields for *S. cerevisiae* and mixed culture at selected cellulase enzyme loadings with and without β-glucosidase supplementation on dilute acid pretreated corn residue crops at 37°C.

	S. cerevisiae							
IU β-glucosidase: IU Cellulose	0:1				8:1			
IU Cellulase/g Cellulose	7	13	19	26	7	13	19	26
Corn Cob	58	63	80	87	87	91	94	94
Corn Stover	54	59	77	84	82	86	90	92
	*Mixed Culture							
Corn Cob	76	85	89	92	92	93	96	96
Corn Stover	75	84	87	89	86	89	92	92

SAC - Final (7 day) saccharification yields for acid pretreated corn cob and stover at selected cellulase enzyme loadings with and without β-glucosidase supplementation at 45°C.**

IU Cellulase/gm Cellulose	7	13	19	26	7	13	19	26
Corn Cob	55	64	78	86	69	83	90	90
Corn Stover	48	64	77	84	64	80	86	89

* Mixed culture: *Saccharomyces cerevisiae* and *Brettanomyces clausenii.*
**Saccharifications are expressed in percent of theoretical conversion.

Appendix 4

Water Balance

PFD-P101-A201

IN/OUT	STREAM	kg/hr	total
IN	101	34,477	
	211	23,080	
	215	7,622	
	216	18,191	
			83,369
OUT	220	62,902	
	520	19,472	
			82,375
Total for PFD			994

PFD-P101-A202

IN/OUT	STREAM	kg/hr	total
IN	220	62,902	
	219	61,082	
	243	18,005	
	245	13,821	
	230	92,637	
			248,447
OUT	247	28,353	
	246	91,676	
	303	12,194	
	403	242	
	410	4,597	
	302	111,360	
			248,423
Total for PFD			25

PFD-P101-A203

IN/OUT	STREAM	kg/hr	total
IN	246	91,676	
			91,676
OUT	230	92,637	
	229	207	
			92,844
Total for PFD			(1,167)

PFD-P101-A301

IN/OUT	STREAM	kg/hr	total
IN	303	12,194	
	421	1,623	
			13,817
OUT	304c	8	
	304	13,781	
			13,789
Total for PFD			28

PFD-P101-A302

IN/OUT	STREAM	kg/hr	total
IN	304	13,781	
	302	121,909	
	551	6,640	
			142,329
OUT	308	141	
	502	139,868	
			140,009
Total for PFD			2,320

PFD-P101-A307

IN/OUT	STREAM	kg/hr	total
IN	302A	111,360	
	307A	10,149	
	422	10,548	
			132,057
OUT	302B	121,909	
	307B	10,149	
			132,057
Total for PFD			-

PFD-P101-A401

IN/OUT	STREAM	kg/hr	total
IN	403	242	
	430	783	
			1,025
OUT	433	866	
	435	182	
			1,048
Total for PFD			(23)

PFD-P101-A402

IN/OUT	STREAM	kg/hr	total
IN	410	4,597	
	433	866	
	411	7,777	
			13,240
OUT	419	1,537	
	421	1,623	
	422	10,548	
			13,709
Total for PFD			(469)

PFD-P101-A501

IN	STREAM	kg/hr	total
	501	139,868	
			139,868
OUT	508	11	
	510	13,909	
	518A	125,948	
			139,868
Total for PFD			-

PFD-P101-A502

IN	STREAM	kg/hr	total
	304c	8	
	308	141	
	508	11	
	524	6,564	
	510	13,909	
	521	879	
			21,512
OUT	550	83	
	551	6,640	
	516	13,919	
	511	924	
			21,565
Total for PFD			(53)

PFD-P101-A503

IN	STREAM	kg/hr	total
	511	924	
			924
OUT	521	879	
	515	45	
			924
Total for PFD			-

PFD-P101-A504

IN	STREAM	kg/hr	total
	518A	125,948	
	610	59,091	
			185,039
OUT	211	22,816	
	243	17,623	
	245	13,664	
	535	10,342	
	531	17,879	
	525	103,980	
			186,303
Total for PFD			(1,264)

PFD-P101-A601

IN	STREAM	kg/hr	total
	604	36,924	
	516	13,919	
	525	103,980	
	531	17,879	
			172,702
OUT	219	61,082	
	411	7,777	
	430	783	
	610	59,091	
	601B	43,969	
			172,702
Total for PFD			-

PFD-P101-A602

IN	STREAM	kg/hr	total
	520	19,472	
	535	10,342	
	494	6,842	
	821	2,699	
	247	28,353	
			67,708
OUT	624	67,708	
			67,708
Total for PFD			-

PFD-P101-A801

IN	STREAM	kg/hr	total
	813	80,536	
			80,536
OUT	815A	12,060	
	215	7,622	
	594	25,190	
	592	3,230	
	237	1,167	
	596	229	
	216	18,191	
	307	10,149	
	821	2,699	
			80,536
Total for PFD			0

PFD-P101-A802

IN	STREAM	kg/hr	total
	815A	12,060	
	811	29,678	
	593	3,230	
	595	25,190	
	307	10,149	
	597	229	
	821	2,699	
			83,235
OUT	813	80,536	
	821	2,699	
			83,235
Total for PFD			-

PFD-P101-A901

IN	STREAM	kg/hr	total
	941	75,268	
	945	5,553,810	
	950	2,276,429	
			7,905,507
OUT	949	64,004	
	940	5,553,810	
	942	4,422	
	944	6,842	
	951	2,276,429	
			7,905,506
Total for PFD			0

PFD-P101-A902

IN	STREAM	kg/hr	total
	904	79,972	
	624	67,708	
			147,680
OUT	524	6,786	
	811	29,678	
	604	36,924	
	906	28	
	941	75,268	
			148,685
Total for PFD			(1,005)

	IN	OUT
Process Totals	9,530,671	9,531,285
NET	(614)	
	-0.01% of in	

Facility Summary
stream numbers

IN	Stream numbers (in)	Stream numbers (out)	OUT
	101	435	
	904	419	
114,449		550	114,449
		620	balance
		949	0
		942	
		229	
		601B	
		515	

Appendix 5

quip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Install Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description	3442 WORK		NREL 900TPD
01	1	0	Bale conveyor	AREA0100	154	170	1.11	$15,000	1999	$15,000	0.6	$15,927	1.5	$24,551	$ 15,927	wire mesh conveyor 60" wide 20' long	WC101	11.93	
02	1	0	Radial Stacker Conveyor	AREA0100	154	170	1.11	$159,830	1999	$159,830	0.6	$169,708	1.5	$261,604	$ 169,708	16 degree, 36" x 200' radial stacker, 750 ton/hr, 75 HP	WC102	44.74	
03	1	0	Breaker Infeed Belt	AREA0100	154	170	1.11	$49,500	1999	$49,500	0.6	$52,559	1.5	$81,020	$ 52,559	84" x 35' rubber belt cleated infeed conveyor, 10 HP, TEFC drive motor with guard	WC103	5.97	
04	1	0	1st Shredder Conveyor	AREA0100	154	170	1.11	$25,650	1999	$25,650	0.6	$27,235	1.5	$41,983	$ 27,235	60" wide x 25' long, 10 HP, TEFC drive with guard	WC104	5.97	
05	1	0	1st Infeed Belt	AREA0100	154	170	1.11	$38,500	1999	$38,500	0.6	$40,879	1.5	$63,015	$ 40,879	60" wide x 30' long, 10 HP, TEFC drive with guard	WC105	11.93	
06	1	0	2nd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.5	$48,285	$ 31,323	48" wide x 20' long, 7.5 HP, TEFC drive with guard	WC106	4.47	
07	1	0	2nd Infeed Belt	AREA0100	154	170	1.11	$27,500	1999	$27,500	0.6	$29,200	1.5	$45,011	$ 29,200	48" wide x 30' long, 5 HP, TEFC drive with guard	WC107	2.98	
08	1	0	3rd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.5	$48,285	$ 31,323	48" wide x 20' long, 10 HP, TEFC drive with guard	WC108	5.97	
09	1	0	Feed Screw Conveyor	AREA0100	225,140	562,850	2.50	$31,700	1997	$31,700	0.6	$54,932	1.5	$86,351	$ 56,018	14" dia. 250' long	WC109	53.75	
01	2	0	Truck Scale	AREA0100	96	72	0.75	$10,000	1999	$20,000	0.6	$16,829	1.5	$25,244	$ 16,829	96 deliveries /scale/12hr			
02	1	0	Receiving Pad	AREA0100	250,000	250,000	1.00	$2,083,500	1999	$2,083,500	0.6	$2,083,500	1.0	$2,083,500	$ 2,083,500	250,000 ft2 concrete pad, 9" thick with drainage			
03	6	1	Front End Loader	AREA0100	159,948	159,948	1.00	$156,000	1998	$1,092,000	0.6	$1,092,000	1.2	$ 1,326,016	$ 1,105,013	run on gasoline			
04	3	0	Bale Breaker	AREA0100	154	170	1.11	$250,000	1999	$750,000	0.6	$796,352	1.2	$955,622	$ 796,352	30 HP each	WM104	53.69	
05	1	0	Primary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.2	$135,444	$ 112,870	250 HP, 1200 rpm, hammermill	WM105	149.14	
06	1	0	Secondary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.5	$169,304	$ 112,870	250 HP, 1200 rpm, hammermill	WM106	149.14	
07	1	0	Shred Bunker	AREA0100	600,000	600,000	1.00	$700,000	1999	$700,000	0.6	$700,000	1.0	$700,000	$ 700,000	200x100x30ft bunker with three walls, 3 days shred storage			
08	1	0	Storm Runoff Pond	AREA0100	1,747,767	1,747,767	1.00	$51,198	1998	$51,198	0.6	$51,198	1.0	$51,198	$ 51,808	200 x 150 x 8 ft, 240,000ft3			295.80
										weighted averages:	0.60		1.13					499.68	295.80
0									Subtotal	$5,315,978		$5,418,705		$6,146,434	$5,433,414				
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.3	$3,181,636	($2,964,798)	is installed cost savings			
										Cost Base Year =		1999							
01	1	0	In-line Sulfuric Acid Mixer	STRM0214	55,308	23,725	0.43	$1,900	1997	$1,900	0.48	$1,266	1.2	$1,585	$1,291	Static Mixer, 110 gpm total flow			
02	1	0	In-line NH3 Mixer	STRM0244	53,630	18,317	0.34	$1,500	1997	$1,500	0.48	$896	1.2	$1,122	$913	Static Mixer, 82 gpm total flow			
09	1	0	Overliming Tank Agitator	STRM0228	167,050	102,608	0.61	$19,800	1997	$19,800	0.51	$15,442	1.2	$19,345	$15,748	Top Mounted, 1800 rpm, 15 hp	WT209	8.39	11.11
24	1	0	Reacidification Tank Agitator	STRM0239	167,280	102,752	0.61	$65,200	1997	$65,200	0.51	$50,851	1.2	$63,702	$51,857	Top-Mounted, 1800 rpm, 54 hp	WT224	25.17	44.49
32	1	0	Reslurrying Tank Agitator	STRM0250	358,810	167,795	0.47	$36,000	1997	$36,000	0.51	$24,432	1.2	$30,606	$24,915	Top-Mounted, 1800 rpm, 25 hp	WT232	13.98	11.99
35	1	0	In-line Acidification Mixer	STRM0236	164,570	101,104	0.61	$2,600	1997	$2,600	0.48	$2,058	1.2	$2,578	$2,099	Static-Mixer, 440 gpm total flow			
01	1	0	Hydrolyzate Screw Conveyor	STRM0220	225,140	101,493	0.45	$59,400	1997	$59,400	0.78	$31,908	1.5	$50,158	$32,539	18" dia. 33' long, 3420 cfh max flow, 23 hp	WC201	13.72	13.83
02	1	0	Wash Solids Screw Conveyor	STRM0225	196,720	165,453	0.84	$23,700	1997	$23,700	1	$19,933	1.5	$31,334	$20,327	18" dia. 16' long, 3420 cfh max flow	WC202	16.70	16.42
25	1	0	Lime Solids Feeder					$3,900	1997	$3,900	1	$3,900	1.5	$6,131	$3,977	6" dia., 63 cfh, 3150 lb/hr max flow	WC225	0.15	0.12
00	1	0	Hydrolyzate Cooler	AREA0200	1,988	895	0.45	$45,000	1997	$45,000	0.51	$29,947	2.2	$66,543	$30,539	Fixed Tube Sheet, 900 sf, 20" dia. X 20' long			
01	1	1	Beer Column Feed Economizer	AREA0201	5,641	5,641	1.00	$139,350	1999	$278,700	0.68	$278,700	2.2	$607,278	$278,700	TEMA type AES shell and tube 5641 sf, 42" dia x 20' long			
02	1	0	Prehydrolysis Reactor	STRM0217	270,034	121,514	0.45	$12,461,841	1998	$12,461,841	0.78	$6,684,746	1.5	$10,146,612	$6,764,408	Vertical Screw, 10 min residence time	WM105	353.16	353.16
01	1	1	Sulfuric Acid Pump	STRM0710	1,647	414	0.25	$4,800	1997	$9,600	0.79	$3,228	2.8	$9,190	$3,291	2 gpm, 245 ft head	WP201	0.40	0.27
09	1	1	Overlimed Hydrolyzate Pump	STRM0228	167,050	102,608	0.61	$10,700	1997	$21,400	0.79	$14,561	2.8	$41,458	$14,849	448 gpm, 150 ft. head	WP209	18.01	24.63
22	1	1	Filtered Hydrolyzate Pump	STRM0230	162,090	101,614	0.63	$10,800	1997	$21,600	0.79	$14,936	2.8	$42,526	$15,231	448 gpm, 150 ft head	WP222	17.83	24.52
23	1	0	Lime Unloading Blower	STRM0227	547	337	0.62	$47,600	1998	$47,600	0.5	$37,340	1.4	$52,898	$37,785	3341 cfm, 6 psi, 10,024 lb/hr	WP223	4.10	4.10
24	1	1	Hydrolysis Feed Pump	STRM0250	160,000	167,795	1.05	$64,934	1999	$129,868	0.6	$133,628	1.2	$160,354	$133,628	740 gpm, 240 ft head	WP224	119.31	27.70
25	1	1	ISEP Elution Pump	STRM0243	52,731	18,005	0.34	$7,900	1997	$15,800	0.79	$6,761	2.8	$19,249	$6,894	104 gpm, 150 ft head	WP225	3.92	3.55
26	1	1	ISEP Reload Pump	STRM0246	164,080	100,802	0.61	$8,700	1997	$17,400	0.79	$11,841	2.8	$33,714	$12,075	445 gpm, 150 ft head	WP226	17.92	12.04
27	1	1	ISEP Hydrolyzate Feed Pump	STRM0221	160,290	98,157	0.61	$10,700	1997	$21,400	0.79	$14,526	2.8	$41,359	$14,814	432 gpm, 150 ft head	WP227	16.81	24.12
39	1	1	Reacidified Liquor Pump	STRM0239	167,280	102,752	0.61	$10,800	1997	$21,600	0.79	$14,698	2.8	$41,847	$14,988	450 gpm, 100 ft head	WP239	12.09	16.47
02	3	0	Pre-IX Belt Filter Press	SOLD0220	57,000	57,000	1.00	$200,000	1998	$600,000	0.39	$600,000	1.4	$850,010	$607,150	Use 3 units for 45% of the flow as recommended by the vendor	WS202	19.69	19.19
21	1	0	ISEP	STRM0240	210,005	98,157	0.47	$2,058,000	1997	$2,058,000	0.33	$1,601,194	1.2	$1,959,422	$1,632,851	10 chambers (39" dia. X 84" high), 4" dia. Valve - Weak Base Resin	WS221	2.98	2.24
22	1	0	Hydroclone & Rotary Drum Filter	STRM0229	5,195	1,137	0.22	$165,000	1998	$165,000	0.39	$91,224	1.4	$129,235	$92,311	Hydrocyclone and Vacuum Filter for 453 gpm	WS222	11.93	11.60
27	1	0	LimeDust Vent Baghouse	STRM0227	548	337	0.61	$32,200	1997	$32,200	1	$19,778	1.5	$30,254	$20,169	3750 cfm, 625 sf, 6 cfm/sf			
01	1	0	Sulfuric Acid Storage	STRM0710	1,647	860	0.52	$5,760	1996	$5,760	0.71	$3,633	1.7	$6,283	$3,751	2000 gal., 24 hr. residence time, 90% wv, 5.5ft diam. X 11ft			
03	1	0	Blowdown Tank	STRM0217	270,300	121,514	0.45	$64,100	1997	$64,100	0.93	$30,475	1.7	$52,061	$31,078	7000 gal., 11' dia x 30' high, 10 min. res. time, 75% wv, 15 psig			
09	1	0	Overliming Tank	STRM0228	167,050	102,608	0.61	$71,000	1997	$71,000	0.71	$50,232	1.8	$90,186	$51,225	29850 gal., 16' dia. X 32' high, 1 hr. res. time, 90% wv, 15 psig			
20	1	0	Lime Storage Bin	STRM0227	548	548	1.00	$69,200	1997	$69,200	0.46	$69,200	1.8	$124,243	$70,568	4455 cf, 14' dia x 25' high, 1.5x rail car vol., atmospheric, 15 day storage max			
24	1	0	Reacidification Tank	STRM0239	102,752	102,752	1.00	$111,889	1999	$111,889	0.51	$111,889	1.8	$196,992	$111,889	120,000 gal., 28' dia x 28' high, 4 hr. res. time, 90% wv, atmospheric			
12	1	0	Slurrying Tank	STRM0250	358,810	167,795	0.47	$44,800	1997	$44,800	0.71	$26,117	1.8	$46,890	$26,633	11300 gal., 13' dia. X 25' high, 15 min. res. time, 90% wv			
									1999						$0				
										weighted averages:	0.70		1.48					676.27	621.55
0									Subtotal	$16,527,758		$9,999,337		$14,955,166	$10,128,493				
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.5	$15,025,380	$70,213	is installed cost savings			

Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Install Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description	3442 WORK		NREL 800TPD
00	8	0	Fermentor Agitators	GALLONS	962,651	750,000	0.78	$19,676	1996	$157,408	0.51	$138,592	1.2	$175,799	$143,110	Side Mounted, 2 per vessel, 60 hp each, 0.15 hp/1000 gal	WT300	201.34	354.50
01	1	0	Seed Hold Tank Agitator	STRM0304	41,777	17,529	0.42	$12,551	1996	$12,551	0.51	$8,060	1.2	$10,223	$8,322	Top Mounted, 1800 rpm, 10 hp, 0.1 hp/1000 gal	WT301	5.59	9.44
04	2	0	4th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$11,700	1997	$23,400	0.51	$15,026	1.2	$18,824	$15,323	Top Mounted, 1800 rpm, 3 hp, 0.3 hp/1000 gal	WT304	3.36	4.72
05	2	0	5th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$10,340	1996	$20,680	0.51	$13,280	1.2	$16,845	$13,713	Top Mounted, 1800 rpm, 9 hp, 0.1 hp/1000 gal	WT305	10.07	15.73
06	1	0	Beer Well Agitator	STRM0502	381,700	173,737	0.46	$10,100	1997	$10,100	0.51	$6,761	1.2	$8,469	$6,894	Top Mounted, 1800 rpm, 2 hp, 0.3 hp/1000 gal	WT306	1.12	1.21
00	4	0	Fermentors	GALLONS	750,000	750,000	1.00	$326,203	1999	$1,304,812	0.71	$1,304,812	1.8	$2,297,260	$1,304,812	750,000 gal. each, 2 day residence total, 90% wv, API, atmospheric, 50' ϕ x 51'			
01	2	0	1st Fermentation Seed Fermentor	None			0.45	$14,700	1997	$29,400	0.93	$13,991	2.8	$39,948	$14,267	9 gal, jacketed, agitated, 1' dia., 1.5' high, 15 psig			
02	2	0	2nd Fermentation Seed Fermentor	None			0.45	$32,600	1997	$65,200	0.93	$31,027	2.8	$88,592	$31,640	90 gal., jacketed, agitated, 2' 3" dia., 3' high, 2.5 psig			
03	2	0	3rd Fermentation Seed Fermentor	None			0.45	$81,100	1997	$162,200	0.93	$77,186	2.8	$220,394	$78,712	900 gal., jacketed, agitated, 5' dia, 6.5' high, 2.5 psig			
04	2	0	4th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$39,500	1997	$79,000	0.93	$35,225	1.7	$60,174	$35,921	9000 gal., 9' dia x 19' high, atmospheric			
05	2	0	5th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$147,245	1998	$294,490	0.51	$189,107	1.8	$336,910	$191,360	90000 gal., API, atmospheric 25'ϕ x 25'			
00	4	1	Fermentation Cooler	QHX300EA	67,820	25,053	0.37	$4,000	1997	$20,000	0.78	$9,198	2.2	$20,438	$9,380	4 exchangers at 221 sf, U=300 BTU/hr sf F LMTD = 22.9°F plate and frame			
01	1	0	Fermentation Seed Hydrolyzate Cooler	AREA0301	773	318	0.41	$15,539	1998	$15,539	0.78	$7,778	2.2	$17,151	$7,871	348 sf, 300 BTU/hr sf F			
02	1	0	Fermentation Pre-Cooler	AREA0302	3,765	828	0.22	$25,409	1998	$25,409	0.78	$7,797	2.2	$17,193	$7,890	828 sf total, plate and frame			
04	1	0	4TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$3,300	1997	$3,300	0.83	$1,580	1.2	$1,934	$1,611	12 sf, 1" sch 40 pipe, 105 BTU/hr sf F			
05	1	0	5TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$18,800	1997	$18,800	0.98	$7,881	1.2	$9,644	$8,037	138 sf, 2" sch 40 pipe, 92 BTU/hr sf F			
00	4	1	Fermentation Recirc./Transfer Pump	QHX300EA	67,737	55,505	0.82	$8,000	1997	$40,000	0.79	$34,177	2.8	$97,307	$34,852	844 gpm @ 150 ft sized based on heating rate	WP300	104.49	277.00
01	1	1	Fermentation Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$22,194	1998	$44,388	0.7	$24,168	1.4	$34,238	$24,456	280 gpm @ 150 ft head	WP301	5.95	6.92
02	2	0	Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$54,088	1998	$108,176	0.7	$58,898	1.4	$83,440	$59,600	504 gpm total, 252 gpm each, 100 ft head	WP302	7.14	6.92
06	1	1	Beer Transfer Pump	STRM0502	381,701	173,737	0.46	$17,300	1997	$34,600	0.79	$18,579	2.8	$52,899	$18,947	790 gpm each, 171 ft head	WP306	34.47	45.74
01	1	0	Fermentation Seed Hold Tank	STRM0304	41,777	17,529	0.42	$161,593	1998	$161,593	0.51	$103,767	1.8	$184,870	$105,003	105000 gal., API atmospheric			
06	1	0	Beer Well	STRM0502	129,000	183,467	1.42	$111,889	1999	$111,889	0.51	$133,906	1.8	$235,756	$133,906	192,518 gal., 32' dia x 32' high, 4 hr. res. time, 95% wv, atmospheric			
										weighted averages:	**0.66**		**1.79**					373.53	722.18
00									**Subtotal**	**$2,742,935**		**$2,240,795**		**$4,028,307**	→ **$2,255,629**				
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.3	$8,218,509	→ $4,190,202	is installed cost savings			
07	8	0	Enzymatic Hydrolysis Tank Agitators	STRM0302B	157,136	157,136	1.00	$19,676	1996	$157,408	0.51	$157,408	1.2	$199,666	$162,539	two side mounted 75 hp agitators / tank, 0.4hp/1000 gal.	WT307	251.67	
07	12	0	Enzymatic Hydrolysis Tank Heater	STRM0302B	157,136	157,136	1.00	$15,000	1999	$180,000	0.78	$180,000	2.2	$392,214	$180,000	65 ft2 double pipe			
08	1	0	Pre-hydrolyzate cooler	STRM0302	145,536	145,536	1.00	$25,000	1999	$25,000	0.78	$25,000	2.2	$54,474	$25,000	481 ft2, parallel double pipe			
08	8	1	Hydrolyzer Bottoms Pump	STRM0302B	157,136	157,136	1.00	$121,690	1999	$1,095,210	0.6	$1,095,210	1.2	$1,314,252	$1,095,210	3000 GPM each Disc flow pumps, 245ft head	WP308	1,744.94	
07	4	0	Enzymatic Hydrolysis Tank	STRM0302B	750,000	375,000	0.50	$326,203	1999	$1,304,812	0.6	$860,855	2.0	$1,753,728	$860,855	~~375,000 gallons, 24 hour residence time, 2 side mounted agitators cone bottom,~~ concrete base, bottom outlet through the concrete, 30° cone bottom			
		0							1999						$0				
										weighted averages:	**0.61**		**1.60**					1,996.61	-
a 307									**Subtotal**	**$2,762,430**		**$2,318,473**		**$3,714,334**	→ **$2,323,604**				
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)				$0	→ ($3,714,334)	is installed cost savings			
															$475,868	**Cost Savings with SHCF (sum of A300 & A307 savings)**			
00	11	0	Cellulase Fermentor Agitators	GALLONS	150,000	88,335	0.59	$200,000	1999	$2,200,000	0.51	$1,679,359	1.2	$2,062,956	$1,679,359	125 hp / agitator -- 1 agitator/vessel	WT400	559.28	1,373.10
00	11	0	Cellulase Fermentors	GALLONS	88,335	88,335	1.00	$179,952	1998	$1,979,472	0.71	$1,979,472	1.8	$3,526,602	$2,003,061	88335 gal, 2.5 psig, cooling coils in tank costed as H400, 40 ft. height, 20 ft. diameter			
01	3	0	1st Cellulase Seed Fermentor	STRM0433	2,790	932	0.33	$22,500	1997	$67,500	0.93	$24,343	2.0	$49,648	$24,824	11 gal / 15 psig / Jacketed / Agitator			
02	3	0	2nd Cellulase Seed Fermentor	STRM0433	2,790	932	0.33	$54,100	1997	$162,300	0.93	$58,531	2.0	$119,377	$59,689	221 gal / 15 psig / Jacketed./Agitator	WT402	119.82	149.78
03	3	0	3rd Cellulase Seed Fermentor	STRM0433	2,790	932	0.33	$282,100	1997	$846,300	0.93	$305,207	2.0	$622,482	$311,241	4417 gal / 15 psig / Jacketed./Agitator			
00	11	0	Cellulase Fermentation Cooler	QHX400EA	236,668	88,335	0.37	$34,400	1997	$378,400	0.78	$175,431	2.2	$389,815	$178,899	Immersible Coil 205 ft2 each			
01	5	1	Fermentor Air Compressor Package	STRM0440	80,455	80,455	1.00	$229,000	1999	$1,374,000	0.34	$1,374,000	1.3	$1,786,200	$1,374,000	7946 scfm each, 50 psig outlet, 1277 hp each, includes starter	WM401	5,108.00	5,370.92
00	1	1	Cellulase Transfer Pump	STRM0420	40,543	11,600	0.29	$9,300	1997	$18,600	0.79	$6,921	2.8	$19,706	$7,058	58 GPM / 100 ft. head	WP400	1.57	2.22
01	1	1	Cellulase Seed Pump	STRM0433	2,790	932	0.33	$12,105	1998	$24,210	0.7	$11,236	1.2	$13,644	$11,370	24 gpm / 1 hp	WP401	0.28	0.31
05	1	1	Media Pump	STRM0416	586	200	0.34	$8,300	1997	$16,600	0.79	$7,104	2.8	$20,227	$7,245	21 Gpm/100 Ft Head	WP405	0.09	0.03
20	1	1	Anti-foam Pump	STRM0417	227	79	0.35	$5,500	1997	$11,000	0.79	$4,761	2.8	$13,555	$4,855	4 gpm / 75 ft head	WP420	0.01	0.01
05	1	0	Media-Prep Tank	STRM0416	586	200	0.34	$64,600	1997	$64,600	0.71	$30,128	1.7	$51,467	$30,723	2083 Gal / 1.17 hp Agitator	WT402	0.65	85.84
20	1	0	Anti-foam Tank	STRM0417	227	79	0.35	$402	1998	$402	0.71	$189	1.7	$321	$192	67 gal, 3 hr. residence time			
										weighted averages:	**0.61**		**1.52**					5,789.71	6,982.21
00									**Subtotal**	**$7,143,384**		**$5,656,682**		**$8,676,000**	**$5,692,516**				
												from Pfpve.xls/0.45 equip.		$10,353,995	$1,677,994	**Installed Cost Savings Using PureVision Enzyme Production Technology**			

Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Install Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description	3442 WORK		NREL 900TPD
501	1	0	Beer Column	DIAM0501	4	2.29	0.56	$636,976	1996	$636,976	0.78	$402,792	2.1	$873,434	$415,921	7'6" DIA, 32 ACTUAL TRAYS, NUTTER V-GRID TRAYS			
502	1	0	Rectification Column	S510S521	56,477	26,744	0.47	$525,800	1996	$525,800	0.78	$293,491	2.1	$636,421	$303,058	8' dia.(rect)., 4' dia. (strip) x 18" T.S., 60 act. Trays, 60% eff., Nutter V-Grid trays			
501	1	0	1st Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,676	1996	$435,676	0.68	$435,676	2.1	$944,742	$449,877	22278 sf each., 135 BTU/hr sf F			
502	1	0	2nd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.1	$944,685	$449,850	22278 sf., 170 BTU/hr sf F			
503	1	0	3rd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.1	$944,685	$449,850	22278 sf each., 170 BTU/hr sf F			
501	1	0	Beer Column Reboiler	QRFD0501	-7,863,670	-3,723,722	0.474	$158,374	1996	$158,374	0.68	$95,263	2.2	$214,340	$98,368	Fixed TS, 6602 sf, 31" dia., 20' long, 178 BTU/hr sf F			
502	1	0	Rectification Column Reboiler	QRFD0502	-987,427	-467,581	0.474	$29,600	1997	$29,600	0.68	$17,805	2.2	$39,563	$18,157	Thermosyphon, 512 sf, 15" dia., 20' long, 130 BTU/hr sf F			
504	1	0	Beer Column Condenser	QCND0501	277,820	131,557	0.474	$29,544	1996	$29,544	0.68	$17,771	2.2	$39,984	$18,350	Floating Head, 418 sf, 15" dia., 22' long, 92 BTU/hr sf F			
505	1	0	Rectification Column Condenser	QCND0502	4,905,410	2,322,883	0.474	$86,174	1996	$86,174	0.68	$51,834	2.2	$116,626	$53,524	Fixed TS, 1969 sf, 29" dia., 20' long, 157 BTU/hr sf F			
512	1	1	Beer Column Feed Interchange	AREA0512	909	430	0.474	$19,040	1996	$38,080	0.68	$22,905	2.2	$51,537	$23,652	431 sf, 200 BTU/hr sf F			
517	1	1	Evaporator Condenser	QHET0517	6,764,222	3,203,095	0.47	$121,576	1996	$243,152	0.68	$146,257	2.2	$329,077	$151,024	Fixed TS, 3906 sf, 29" dia., 20' long, 220 BTU/hr sf F			
503	1	0	Molecular Sieve (9 pieces)	STRM0515	20,491	9,703	0.47	$2,700,000	1998	$2,700,000	0.7	$1,599,964	1.0	$1,619,030	$1,619,030	Superheater, twin mole sieve columns, product cooler, condenser, pumps, vacuum source.	WM503	55.00	55.00
501	1	1	Beer Column Bottoms Pump	P501FLOW	5,053	2,200	0.44	$42,300	1997	$84,600	0.79	$43,861	2.8	$124,881	$44,728	2200 gpm, 150 ft head	WP501	84.65	118.68
503	1	1	Beer Column Reflux Pump	QCND0501	277,820	131,557	0.47	$1,357	1998	$2,714	0.79	$1,504	2.8	$4,248	$1,522	6 gpm, 140 ft head	WP503	0.22	0.51
504	1	1	Rectification Column Bottoms Pump	STRM0516	31,507	15,530	0.49	$4,916	1998	$9,832	0.79	$5,622	2.8	$15,884	$5,689	76 gpm, 158 ft head	WP504	2.80	3.46
505	1	1	Rectification Column Reflux Pump	QCND0502	4,906,301	2,323,304	0.47	$4,782	1998	$9,564	0.79	$5,299	2.8	$14,970	$5,362	207 gpm, 110 ft head	WP505	5.14	12.77
511	2	1	1st Effect Pump	STRM0525	278,645	133,617	0.48	$19,700	1997	$59,100	0.79	$33,069	2.8	$94,155	$33,723	1137 gpm each, 110 ft head	WP511	67.89	80.57
512	1	1	2nd Effect Pump	STRM0528	91,111	45,390	0.50	$13,900	1997	$27,800	0.79	$16,032	2.8	$45,646	$16,349	599 gpm, 110 ft head	WP512	17.37	19.12
513	2	1	3rd Effect Pump	STRM0531	48,001	23,814	0.50	$8,000	1997	$24,000	0.79	$13,795	2.8	$39,276	$14,068	196 gpm each, 110 ft head	WP513	12.54	10.26
514	1	1	Evaporator Condensate Pump	STRM534A	140,220	69,285	0.49	$12,300	1997	$24,600	0.79	$14,095	2.8	$40,131	$14,374	293 gpm, 125 ft head	WP514	9.20	12.43
515	1	1	Scrubber Bottoms Pump	STRM0551	15,377	7,427	0.48	$2,793	1998	$5,586	0.79	$3,143	2.8	$8,881	$3,181	31 gpm, 104 ft head	WP515	0.84	0.77
517	1	1	Kill Tank Bottoms Pump	STRM0518	5,053	660	0.13	$42,300	1997	$84,600	0.79	$16,944	2.8	$48,242	$17,279	660gpm, 72 ft head	WP517	12.19	-
503	1	0	Beer Column Relfux Drum	QCND0501	277,820	131,557	0.47	$11,900	1997	$11,900	0.93	$5,938	1.7	$10,144	$6,055	164 gal, 15 min res. Time, 50% wv, 2'6" dia., 5' long, 25 psig			
505	1	0	Rectification Column Reflux Drum	QCND0502	4,906,301	2,323,304	0.47	$45,600	1997	$45,600	0.72	$26,621	1.7	$45,476	$27,147	6225 gal, 15 min res time, 50% wv, 7' dia, 22' long, 25 psig			
512	1	0	Vent Scrubber	STRM0523	18,523	9,768	0.53	$99,000	1998	$99,000	0.78	$60,197	1.7	$102,043	$60,915	5' dia x 25' high, 4 stages, plastic Jaeger Tri-Packing			
513	1	0	Kill Tank	STRM0518	149,897	149,897	1.00	$99,920	1999	$99,920	0.78	$99,920	1.7	$167,384	$99,920	18 psig, 30 min. res. time			
										weighted averages:	0.72		1.71					267.85	313.57
00									Subtotal	$6,343,492		$4,301,097		$7,515,486	$4,400,972				
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.7	$6,765,614	($749,872)	is installed cost savings			
501	1	0	Lignin conveyor	STRM0601B	225,140	225,140	1.00	$31,700	1997	$31,700	0.6	$31,700	1.5	$49,832	$ 32,327	14" dia, 100' long	WC109	21.50	
613	1	0	Syrup Sprayer	STRM0531	22,372	22,372	1.00	$1,000	1999	$1,000	0.3	$1,000	1.2	$1,200	$1,000	100 GPM syrup sprayer			
614	1	0	Lignin Loadout	STRM0601A	63,778	0	0.00	$41,200	1999	$41,200	0.3	$0	1.0	$0	$0	245 GPM @ 20.6% insoluble solids			
615	1	0	Equalization Basin	STRM0830	98,267	102,204	1.04	$350,000	1999	$350,000	0.79	$361,031	1.0	$361,031	$361,031	no less than 500,000 gal., above-ground bolted tank with cover, including foundations, pumps and controls	WM615	1,077.21	-
616	1	0	Anaerobic Digestion System	STRM0830	98,267	102,204	1.04	$3,200,000	1999	$3,200,000	0.79	$3,300,852	1.0	$3,300,852	$3,300,852	500,000 gal., includes site work, foundations, reactors and ancillary equipment			
617	1	0	Aerobic Digestion System	STRM0830	98,267	102,204	1.04	$4,300,000	1999	$4,300,000	0.79	$4,435,520	1.0	$4,435,520	$4,435,520	four-350,000 gal. Sequencing Batch Reactors, 48,000 lbs/day of O_2 transfer capability, de-nitrification facilities, aeration and mixing requires approximately 1,400 horsepower			
618	1	0	Pressure Sand Filters	STRM0830	98,267	102,204	1.04	$280,000	1999	$280,000	0.79	$288,825	1.0	$288,825	$288,825	400 ft^2 of filtration surface area, includes the engineering and legal cost to acquire an NPDES permit			
30	1	1	Recycle Water Pump	STRM0602	179,446	84,120	0.47	$10,600	1997	$21,200	0.79	$11,652	2.8	$33,175	$11,882	370 gpm, 150ft head	WP630	14.75	-
01	2	0	Beer Column Bottoms Centrifuge	CENTFLOW	404	300	0.74	$659,550	1998	$1,319,100	0.6	$1,103,371	1.2	$1,339,824	$1,116,520	requires 540gpm duty, 2 @ 300 gpm and 410 hp each	WS601	489.18	400.03
30	1	0	Recycled Water Tank	STRM0602	179,446	84,120	0.47	$14,515	1998	$14,515	0.745	$8,254	1.7	$13,992	$8,353	7410 gal, 20 min. res., 2.5 psig, 9.5ft diam. x 14.25ft			
										weighted averages:	0.76		1.03					1,602.64	690.39
00									Subtotal	$9,558,715		$9,542,206		$9,824,251	$9,556,310				
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.3	$5,167,342	($4,656,910)	is installed cost savings			

Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost In Base Year	Install Factor	Installed Cost	Scaled Uninstalled Cost In 1999$	Description	3442 WORK		NREL 900TPD
03	1	1	Sulfuric Acid Pump	STRM0710	1,647	1,912	1.16	$8,000	1997	$16,000	0.79	$18,001	2.8	$51,253	$18,357	215 gpm, 150ft head	WP703	0.09	0.09
07	1	1	Antifoam Store Pump	STRM0417	227	79	0.35	$5,700	1997	$11,400	0.79	$4,934	2.8	$14,048	$5,031	0.5 gpm, 92 ft head	WP707	0.01	0.01
20	1	1	CSL Pump	STRM0735	2,039	859	0.42	$8,800	1997	$17,600	0.79	$8,889	2.8	$25,308	$9,065	182 gpm, 150ft head	WP720	0.15	0.18
03	1	0	Sulfuric Acid Storage Tank	STRM0710	1,647	1,912	1.16	$42,500	1997	$42,500	0.51	$45,860	1.8	$82,338	$46,767	20,000 gal, 240 hr supply, 90% wv, 12ft diam. x 24 ft, atmospheric			
07	1	0	Antifoam Storage Tank	STRM0417	227	227	1.00	$14,400	1997	$14,400	0.71	$14,400	1.7	$24,600	$14,685	12,000 gal, 27 day supply, 10.5ft diam. X 18.5ft			
20	1	0	CSL Storage Tank	STRM0735	2,039	859	0.42	$88,100	1997	$88,100	0.79	$44,495	1.7	$76,011	$45,375	30160 gal, 90% wv, 120 supply, 14.3ft diam. X 25 ft			
										weighted averages:	0.72		1.96					0.25	0.28
00									Subtotal	$190,000		$136,579		$273,557	$139,279				
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.5	$1,220,544	$946,987	is installed cost savings			
803	1	0	Boiler with Superheater	STRM0815 + 216	200,000	200,000	1.00	$1,590,000	1999	$1,590,000	0.7	$1,590,000	1.3	$2,067,000	$1,590,000	200,000 #/hr running @ 171,488 #/hr; with 40,000 #/hr 160° superheat; 132,000#/hr 390° sat. @ 205 psig	WM803	75.60	75.60
820	1	0	Hot process water softener system	STRM0811B	229,386	45,003	0.20	$1,383,300	1999	$1,383,300	0.6	$520,623	1.2	$624,748	$520,623	200 gpm			
830	1	0	Hydrazine Addition Pkg.	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM830	10.00	10.00
832	1	0	Ammonia Addition Pkg	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM832	10.00	10.00
834	1	0	Phosphate Addition Pkg.	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM834	10.00	10.00
04	2	1	Condensate Pump	STRM811A	249,633	38,798	0.16	$7,100	1997	$21,300	0.79	$4,894	4.6	$22,958	$4,991	130 gpm, 150' head	WP804	9.21	7.66
24	2	1	Deaerator Feed Pump	STRM811A	196,000	38,798	0.20	$9,500	1997	$28,500	0.79	$7,927	8.3	$67,097	$8,084	180 gpm, 115' head	WP824	4.89	2.27
26	4	1	BFW Pump	STRM0813	207,310	80,536	0.39	$52,501	1998	$262,505	0.79	$124,377	1.4	$176,203	$125,859	310 gpm, 2740' head	WP826	400.99	399.04
28	1	1	Blowdown Pump	STRM0821	6,600	2,699	0.41	$5,100	1997	$10,200	0.79	$5,032	6.4	$32,842	$5,132	12 gpm, 150' head	WP828	0.42	0.93
30	1	1	Hydrazine Transfer Pump	STRM813A	229,386	80,536	0.35	$5,500	1997	$11,000	0.79	$4,811	6.4	$31,402	$4,907	3 gpm, 75' head	WP830	0.05	0.01
04	1	0	Condensate Collection Tank	STRM811A	229,386	38,798	0.17	$7,100	1997	$7,100	0.71	$2,011	3.3	$6,766	$2,050	200 gal, 1.5 min. res. time			
24	1	0	Condensate Surge Drum	STRM811A	150,000	38,798	0.26	$49,600	1997	$49,600	0.72	$18,734	5.0	$95,523	$19,105	2100 gal., 6' diam. X 10', 15 psig, res. time 11 min.			
26	1	0	Deaerator	STRM0813	267,000	80,536	0.30	$165,000	1998	$165,000	0.72	$69,616	6.5	$457,896	$70,446	3030 gal., 15 psig, 10 min. res.			
28	1	0	Blowdown Flash Drum	STRM0821	6,550	2,699	0.41	$9,200	1997	$9,200	0.72	$4,859	7.3	$36,168	$4,955	210 gal., 2.5' diam. X 6', 50 psig 17 min. res.			
30	1	0	Hydrazine Drum	STRM813A	229,386	80,536	0.35	$12,400	1997	$12,400	0.93	$4,685	7.0	$33,440	$4,777	138 gal, 3.75' x 1.25' diam., 10 psig			
										weighted averages:	0.67		1.54					521.16	515.51
00									Subtotal	$3,607,105		$2,387,986		$3,684,612	$2,393,497				
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.1	$23,046,972	$19,362,360	is installed cost savings			
902	1	0	Cooling Tower System	QCWCAPIT	41,100,000	12,955,985	0.32	$1,659,000	1998	$1,659,000	0.78	$674,181	1.2	$818,659	$682,216	40,000 gpm, 185.4MM BTU/hr	WM902	298.85	306.51
904	1	0	Plant Air Compressor	STRM0101	159,950	159,950	1.00	$60,100	1997	$60,100	0.34	$60,100	1.3	$79,675	$61,288	450 cfm, 125 psig outlet	WM904	186.40	186.40
908	1	0	Chilled Water Package	QCHLWCAP	5,040,000	2,268,000	0.45	$380,000	1997	$380,000	0.8	$200,610	1.2	$245,492	$204,577	1000 ton, 600kW	WM908	600.00	507.11
910	1	0	CIP System	STRM0914	63	28	0.45	$95,000	1995	$95,000	0.6	$58,837	1.2	$73,021	$60,851	designed by Delta-T, (est 0.2 kW)	WM910	0.20	-
02	1	1	Cooling Water Pumps	STRM0940	18,290,000	5,553,791	0.30	$332,300	1997	$664,600	0.79	$259,201	2.8	$737,993	$264,326	12300 gpm, 70ft head			
12	1	1	Make-up Water Pump	STRM0904	244,160	82,445	0.34	$10,800	1997	$21,600	0.79	$9,161	2.8	$26,084	$9,343	370 gpm, 75ft head	WP912	7.32	8.00
14	1	1	Process Water Circulating Pump	STRM0905	352,710	111,503	0.32	$11,100	1997	$22,200	0.79	$8,938	2.8	$25,449	$9,115	745 gpm, 75ft head	WP914	14.78	22.38
04	1	1	Instrument Air Dryer	STRM0101	159,950	71,977	0.45	$15,498	1999	$30,996	0.6	$19,197	1.3	$24,956	$19,197	134 scfm air dryer, -40F Dewpoint	WS601	4.91	4.91
04	1	0	Plant Air Receiver	STRM0101	159,950	53,316	0.33	$13,000	1997	$13,000	0.72	$5,894	1.7	$10,069	$6,011	300 gal., 200 psig			
14	1	0	Process Water Tank	STRM0905	352,710	111,503	0.32	$195,500	1997	$195,500	0.51	$108,663	1.8	$195,095	$110,811	234360 gal, 8hr res. time			
										weighted averages:	0.75		1.57			400 gpm well pump, 500ft head	53.16	1,165.62	1,035.31
a 900									Subtotal	$3,141,996		$1,404,783		$2,236,491	$1,427,733		Total kW	12,893	11,177
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.3	$2,895,441	$658,949	is installed cost savings			

3442 PLANT TOTAL:	$57,333,793	$43,406,643	$61,054,640
45% NREL TOTAL:			$75,875,432
SAVINGS:			$14,820,792
			19.53%

)	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
)-P100-A201	A-201	1	0	In-line Sulfuric Acid Mixer	STRM0214	55,308	23,725	0.43	$1,900	1997	$1,900	0.48	$1,266	1.23	$1,585	$1,291	Static Mixer, 110 gpm total flow
)-P100-A202	A-202	1	0	In-line NH3 Mixer	STRM0244	53,630	18,317	0.34	$1,500	1997	$1,500	0.48	$896	1.23	$1,122	$913	Static Mixer, 82 gpm total flow
)-P100-A203	A-209	1	0	Overliming Tank Agitator	STRM0228	167,050	102,608	0.61	$19,800	1997	$19,800	0.51	$15,442	1.23	$19,345	$15,748	Top Mounted, 1800 rpm, 15 hp
)-P100-A203	A-224	1	0	Reacidification Tank Agitator	STRM0239	167,280	102,752	0.61	$65,200	1997	$65,200	0.51	$50,851	1.23	$63,702	$51,857	Top-Mounted, 1800 rpm, 54 hp
)-P100-A202	A-232	1	0	Reslurrying Tank Agitator	STRM0250	358,810	167,795	0.47	$36,000	1997	$36,000	0.51	$24,432	1.23	$30,606	$24,915	Top-Mounted, 1800 rpm, 25 hp
)-P100-A203	A-235	1	0	In-line Acidification Mixer	STRM0236	164,570	101,104	0.61	$2,600	1997	$2,600	0.48	$2,058	1.23	$2,578	$2,099	Static-Mixer, 440 gpm total flow
)-P100-A302	A-300	8	0	Fermentor Agitators	GALLONS	962,651	750,000	0.78	$19,676	1996	$157,408	0.51	$138,592	1.23	$175,799	$143,110	Side Mounted, 2 per vessel, 60 hp each, 0.15 hp/1000 gal
)-P100-A301	A-301	1	0	Seed Hold Tank Agitator	STRM0304	41,777	17,529	0.42	$12,551	1996	$12,551	0.51	$8,060	1.23	$10,223	$8,322	Top Mounted, 1800 rpm, 10 hp, 0.1 hp/1000 gal
)-P100-A301	A-304	2	0	4th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$11,700	1997	$23,400	0.51	$15,026	1.23	$18,824	$15,323	Top Mounted, 1800 rpm, 3 hp, 0.3 hp/1000 gal
)-P100-A301	A-305	2	0	5th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$10,340	1996	$20,680	0.51	$13,280	1.23	$16,845	$13,713	Top Mounted, 1800 rpm, 9 hp, 0.1 hp/1000 gal
)-P100-A302	A-306	1	0	Beer Well Agitator	STRM0502	381,700	173,737	0.46	$10,100	1997	$10,100	0.51	$6,761	1.23	$8,469	$6,894	Top Mounted, 1800 rpm, 2 hp, 0.3 hp/1000 gal
)-P100-A307	A-307	8	0	Enzymatic Hydrolysis Tank Agitators	STRM0302B	157,136	157,136	1.00	$19,676	1996	$157,408	0.51	$157,408	1.23	$199,666	$162,539	two side mounted 75 hp agitators / tank, 0.4hp/1000 gal.
)-P100-A402	A-400	11	0	Cellulase Fermentor Agitators	GALLONS	150,000	88,335	0.59	$200,000	1999	$2,200,000	0.51	$1,679,359	1.23	$2,062,956	$1,679,359	125 hp / agitator -- 1 agitator/vessel
		39	**0**	**39**										**1.23**	**$ 2,611,720**	**$ 66,967**	
		sum	sum	total										avg.	sum	avg. (installed)	

)	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
)-P100-A101	C-101	1	0	Bale conveyor	AREA0100	154	170	1.11	$15,000	1999	$15,000	0.6	$15,927	1.54	$24,551	$ 15,927	wire mesh conveyor 60" wide 20' long
)-P100-A101	C-102	1	0	Radial Stacker Conveyor	AREA0100	154	170	1.11	$159,830	1999	$159,830	0.6	$169,708	1.54	$261,604	$ 169,708	16 degree, 36" x 200' radial stacker, 750 ton/hr, 75 HP
)-P100-A101	C-103	1	0	Breaker Infeed Belt	AREA0100	154	170	1.11	$49,500	1999	$49,500	0.6	$52,559	1.54	$81,020	$ 52,559	84" x 35' rubber belt cleated infeed conveyor, 10 HP, TEFC drive motor with guard
)-P100-A101	C-104	1	0	1st Shredder Conveyor	AREA0100	154	170	1.11	$25,650	1999	$25,650	0.6	$27,235	1.54	$41,983	$ 27,235	60" wide x 25' long, 10 HP, TEFC drive with guard
)-P100-A101	C-105	1	0	1st Infeed Belt	AREA0100	154	170	1.11	$38,500	1999	$38,500	0.6	$40,879	1.54	$63,015	$ 40,879	60" wide x 30' long, 10 HP, TEFC drive with guard
)-P100-A101	C-106	1	0	2nd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.54	$48,285	$ 31,323	48" wide x 20' long, 7.5 HP, TEFC drive with guard
)-P100-A101	C-107	1	0	2nd Infeed Belt	AREA0100	154	170	1.11	$27,500	1999	$27,500	0.6	$29,200	1.54	$45,011	$ 29,200	48" wide x 30' long, 5 HP, TEFC drive with guard
)-P100-A101	C-108	1	0	3rd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.54	$48,285	$ 31,323	48" wide x 20' long, 10 HP, TEFC drive with guard
-P100-A101	C-109	1	0	Feed Screw Conveyor	AREA0100	225,140	562,850	2.50	$31,700	1997	$31,700	0.6	$54,932	1.54	$86,351	$ 56,018	14" dia. 250' long
-P100-A201	C-201	1	0	Hydrolyzate Screw Conveyor	STRM0220	225,140	101,493	0.45	$59,400	1997	$59,400	0.78	$31,908	1.54	$50,158	$32,539	18" dia. 33' long, 3420 cfh max flow, 23 hp
-P100-A202	C-202	1	0	Wash Solids Screw Conveyor	STRM0225	196,720	165,453	0.84	$23,700	1997	$23,700	1.00	$19,933	1.54	$31,334	$20,327	18" dia. 16' long, 3420 cfh max flow
-P100-A203	C-225	1	0	Lime Solids Feeder	none				$3,900	1997	$3,900	1	$3,900	1.54	$6,131	$3,977	6" dia., 63 cfh, 3150 lb/hr max flow
-P100-A601	C-601	1	0	Lignin conveyor	STRM0601B	225,140	225,140	1.00	$31,700	1997	$31,700	0.6	$31,700	1.54	$49,832	$32,327	14" dia. 100' long
		13	**0**	**13**										**1.54**	**$ 837,560**	**$ 64,428**	
		sum	sum	total										avg.	sum	avg. (installed)	

)	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
-P100-A501	D-501	1	0	Beer Column	DIAMD501	4	2.29	0.56	$636,976	1996	$636,976	0.78	$402,792	2.10	$873,434	$415,921	7'6" DIA, 32 ACTUAL TRAYS, NUTTER V-GRID TRAYS
-P100-A502	D-502	1	0	Rectification Column	S510S521	56,477	26,744	0.47	$525,800	1996	$525,800	0.78	$293,491	2.10	$636,421	$303,058	8' dia.(rect)., 4' dia. (strip) x 18" T.S., 60 act. Trays, 60% eff., Nutter V-Grid trays
-P100-A504	E-501	1	0	1st Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,676	1996	$435,676	0.68	$435,676	2.10	$944,742	$449,877	22278 sf each., 135 BTU/hr sf F
-P100-A504	E-502	1	0	2nd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.10	$944,685	$449,850	22278 sf., 170 BTU/hr sf F
-P100-A504	E-503	1	0	3rd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.10	$944,685	$449,850	22278 sf each., 170 BTU/hr sf F
		5	**0**	**5**										**2.10**	**$ 4,343,968**	**$ 868,794**	
		sum	sum	total										avg.	sum	avg. (installed)	

)	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
-P100-A302	F-300	4	0	Fermentors	GALLONS	750,000	750,000	1.00	$326,203	1999	$1,304,812	0.71	$1,304,812	1.76	$2,297,260	$1,304,812	750,000 gal. each, 2 day residence total, 90% wv, API, atmospheric, 50' f x 51'
-P100-A301	F-301	2	0	1st Fermentation Seed Fermentor	None		0	0.45	$14,700	1997	$29,400	0.93	$13,991	2.80	$39,948	$14,267	9 gal, jacketed, agitated, 1' dia., 1.5' high, 15 psig
-P100-A301	F-302	2	0	2nd Fermentation Seed Fermentor	None		0	0.45	$32,600	1997	$65,200	0.93	$31,027	2.80	$88,592	$31,640	90 gal., jacketed, agitated, 2' 3" dia., 3' high, 2.5 psig
-P100-A301	F-303	2	0	3rd Fermentation Seed Fermentor	None		0	0.45	$81,100	1997	$162,200	0.93	$77,186	2.80	$220,394	$78,712	900 gal., jacketed, agitated, 5' dia, 6.5' high, 2.5 psig
-P100-A301	F-304	2	0	4th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$39,500	1997	$79,000	0.93	$35,225	1.68	$60,174	$35,921	9000 gal., 9' dia x 19' high, atmospheric
-P100-A301	F-305	2	0	5th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$147,245	1998	$294,490	0.51	$189,107	1.76	$336,910	$191,360	90000 gal., API, atmospheric 25'f x 25'
-P100-A402	F-400	11	0	Cellulase Fermentors	GALLONS	88,335	88,335	1.00	$179,952	1998	$1,979,472	0.71	$1,979,472	1.76	$3,526,602	$2,003,061	88335 gal, 2.5 psig, cooling coils in tank costed as H400, 40 ft. height, 20 ft. diameter
-P100-A401	F-401	3	0	1st Cellulase Seed Fermentor	STRM0433	2,790	932	0.33	$22,500	1997	$67,500	0.93	$24,343	2.00	$49,648	$24,824	11 gal / 15 psig / Jacketed / Agitator
-P100-A401	F-402	3	0	2nd Cellulase Seed Fermentor	STRM0433	2,790	932	0.33	$54,100	1997	$162,300	0.93	$58,531	2.00	$119,377	$59,689	221 gal / 15 psig / Jacketed./Agitator
-P100-A401	F-403	3	0	3rd Cellulase Seed Fermentor	STRM0433	2,790	932	0.33	$282,100	1997	$846,300	0.93	$305,207	2.00	$622,482	$311,241	4417 gal / 15 psig / Jacketed./Agitator
		34	**0**	**34**										**2.14**	**$ 7,361,387**	**$ 216,511**	
		sum	sum	total										avg.	sum	avg. (installed)	

D	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
D-P100-A202	H-200	1	0	Hydrolyzate Cooler	AREA0200	1,988	895	0.45	$45,000	1997	$45,000	0.51	$29,947	2.18	$66,543	$30,539	Fixed Tube Sheet, 900 sf, 20" dia. X 20' long
D-P100-A201	H-201	1	1	Beer Column Feed Economizer	AREA0201	5,641	5,641	1.00	$139,350	1999	$278,700	0.68	$278,700	2.18	$607,278	$278,700	TEMA type AES shell and tube 5641 sf, 42" dia x 20' long
D-P100-A302	H-300	4	1	Fermentation Cooler	QHX300EA	67,820	25,053	0.37	$4,000	1997	$20,000	0.78	$9,198	2.18	$20,438	$9,380	4 exchangers at 221 sf, U=300 BTU/hr sf F LMTD = 22.9°F plate and frame
D-P100-A301	H-301	1	0	Fermentation Seed Hydrolyzate Cooler	AREA0301	773	318	0.41	$15,539	1998	$15,539	0.78	$7,778	2.18	$17,151	$7,871	348 sf, 300 BTU/hr sf F
D-P100-A302	H-302	1	0	Fermentation Pre-Cooler	AREA0302	3,765	828	0.22	$25,409	1998	$25,409	0.78	$7,797	2.18	$17,193	$7,890	828 sf total, plate and frame
D-P100-A301	H-304	1	0	4TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$3,300	1997	$3,300	0.83	$1,580	1.20	$1,934	$1,611	12 sf, 1" sch 40 pipe, 105 BTU/hr sf F
D-P100-A301	H-305	1	0	5TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$18,800	1997	$18,800	0.98	$7,881	1.20	$9,644	$8,037	138 sf, 2" sch 40 pipe, 92 BTU/hr sf F
D-P100-A307	H-307	12	0	Enzymatic Hydrolysis Tank Heater	STRM0302B	157,136	157,136	1.00	$15,000	1999	$180,000	0.78	$180,000	2.18	$392,214	$180,000	65 ft2 double pipe
D-P100-A307	H-308	1	0	Pre-hydrolyzate cooler	STRM0302	145,536	145,536	1.00	$25,000	1999	$25,000	0.78	$25,000	2.18	$54,474	$25,000	481 ft2, parallel double pipe
D-P100-A402	H-400	11	0	Cellulase Fermentation Cooler	QHX400EA	236,668	88,335	0.37	$34,400	1997	$378,400	0.78	$175,431	2.18	$389,815	$178,899	Immersible Coil 205 ft2 each
D-P100-A501	H-501	1	0	Beer Column Reboiler	QRFD0501	-7,863,670	-3,723,722	0.474	$158,374	1996	$158,374	0.68	$95,263	2.18	$214,340	$98,368	Fixed TS, 6602 sf, 31" dia., 20' long, 178 BTU/hr sf F
D-P100-A502	H-502	1	0	Rectification Column Reboiler	QRFD0502	-987,427	-467,581	0.474	$29,600	1997	$29,600	0.68	$17,805	2.18	$39,563	$18,157	Thermosyphon, 512 sf, 15" dia., 20' long, 130 BTU/hr sf F
D-P100-A501	H-504	1	0	Beer Column Condenser	QCND0501	277,820	131,557	0.474	$29,544	1996	$29,544	0.68	$17,771	2.18	$39,984	$18,350	Floating Head, 418 sf, 15" dia., 22' long, 92 BTU/hr sf F
D-P100-A502	H-505	1	0	Rectification Column Condenser	QCND0502	4,905,410	2,322,883	0.474	$86,174	1996	$86,174	0.68	$51,834	2.18	$116,626	$53,524	Fixed TS, 1969 sf, 29" dia, 20' long, 157 BTU/hr sf F
D-P100-A501	H-512	1	1	Beer Column Feed Interchange	AREA0512	909	430	0.474	$19,040	1996	$38,080	0.68	$22,905	2.18	$51,537	$23,652	431 sf, 200 BTU/hr sf F
D-P100-A504	H-517	1	1	Evaporator Condenser	QHET0517	6,764,222	3,203,095	0.47	$121,576	1996	$243,152	0.68	$146,257	2.18	$329,077	$151,024	Fixed TS, 3906 sf, 29" dia., 20' long, 220 BTU/hr sf F
		40	**4**	**44**										**2.06**	**$ 2,367,812**	**$ 53,814**	
		sum	sum	total										avg.	sum	avg. (installed)	

D	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
D-P100-A101	M-101	2	0	Truck Scale	AREA0100	96	72	0.75	$10,000	1999	$20,000	0.6	$16,829	1.50	$25,244	$ 16,829	96 deliveries /scale/12hr
D-P100-A101	M-102	1	0	Receiving Pad	AREA0100	250,000	250,000	1.00	$2,083,500	1999	$2,083,500	0.6	$2,083,500	1.00	$2,083,500	$ 2,083,500	250,000 ft2 concrete pad, 9" thick with drainage
D-P100-A101	M-103	6	1	Front End Loader	AREA0100	159,948	159,948	1.00	$156,000	1998	$1,092,000	0.6	$1,092,000	1.20	$ 1,326,016	$ 1,105,013	run on gasoline
D-P100-A101	M-104	3	0	Bale Breaker	AREA0100	154	170	1.11	$250,000	1999	$750,000	0.6	$796,352	1.20	$955,622	$ 796,352	30 HP each
D-P100-A101	M-105	1	0	Primary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.20	$135,444	$ 112,870	250 HP, 1200 rpm, hammermill
D-P100-A101	M-106	1	0	Secondary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.50	$169,304	$ 112,870	250 HP, 1200 rpm, hammermill
D-P100-A101	M-107	1	0	Shred Bunker	AREA0100	600,000	600,000	1.00	$700,000	1999	$700,000	0.6	$700,000	1.00	$700,000	$ 700,000	200x100x30ft bunker with three walls, 3 days shred storage
D-P100-A101	M-108	1	0	Storm Runoff Pond	AREA0100	1,747,767	1,747,767	1.00	$51,198	1998	$51,198	0.6	$51,198	1.00	$51,198	$ 51,808	200 x 150 x 8 ft, 240,000ft3
D-P100-A201	M-202	1	0	Prehydrolysis Reactor	STRM0217	270,034	121,514	0.45	$12,461,841	1998	$12,461,841	0.78	$6,684,746	1.50	$10,146,612	$6,764,408	Vertical Screw, 10 min residence time
D-P100-A402	M-401	5	1	Fermentor Air Compressor Package	STRM0440	80,455	80,455	1.00	$229,000	1999	$1,374,000	0.34	$1,374,000	1.30	$1,786,200	$1,374,000	7946 scfm each, 50 psig outlet, 1277 hp each, includes starter
D-P100-A503	M-503	1	0	Molecular Sieve (9 pieces)	STRM0515	20,491	9,703	0.47	$2,700,000	1998	$2,700,000	0.7	$1,599,964	1.00	$1,619,030	$1,619,030	Superheater, twin mole sieve columns, product cooler, condenser, pumps, vacuum source.
D-P100-A601	M-613	1	0	Syrup Sprayer	STRM0531	22,372	22,372	1.00	$1,000	1999	$1,000	0.3	$1,000	1.20	$1,200	$1,000	100 GPM syrup sprayer
D-P100-A601	M-614	1	0	Lignin Loadout	STRM0601A	63,778	0	0.00	$41,200	1999	$41,200	0.3	$0	1.00	$0	$0	245 GPM @ 20.6% insoluble solids
D-P100-A602	M-615	1	0	Equalization Basin	STRM0830	98,267	102,204	1.04	$350,000	1999	$350,000	0.79	$361,031	1.00	$361,031	$361,031	no less than 500,000 gal., above-ground bolted tank with cover, including foundations, pumps and controls
D-P100-A602	M-616	1	0	Anaerobic Digestion System	STRM0830	98,267	102,204	1.04	$3,200,000	1999	$3,200,000	0.79	$3,300,852	1.00	$3,300,852	$3,300,852	500,000 gal., includes site work, foundations, reactors and ancillary equipment
D-P100-A602	M-617	1	0	Aerobic Digestion System	STRM0830	98,267	102,204	1.04	$4,300,000	1999	$4,300,000	0.79	$4,435,520	1.00	$4,435,520	$4,435,520	four-350,000 gal. Sequencing Batch Reactors, 48,000 lbs/day of O2 transfer capability, de-nitrification facilities, aeration and mixing requires approximately 1,400 horsepower
D-P100-A602	M-618	1	0	Pressure Sand Filters	STRM0830	98,267	102,204	1.04	$280,000	1999	$280,000	0.79	$288,825	1.00	$288,825	$288,825	400 ft2 of filtration surface area, includes the engineering and legal cost to acquire an NPDES permit
D-P100-A801	M-803	1	0	Boiler with Superheater	STRM0815 + 216	200,000	200,000	1.00	$1,590,000	1999	$1,590,000	0.7	$1,590,000	1.30	$2,067,000	$1,590,000	200,000 #/hr running @ 171,488 #/hr; with 40,000 #/hr 160o superheat; 132,000#/hr 390o sat. @ 205 psig
D-P100-A802	M-820	1	0	Hot process water softener system	STRM0811B	229,386	45,003	0.20	$1,383,300	1999	$1,383,300	0.6	$520,623	1.20	$624,748	$520,623	200 gpm
D-P100-A803	M-830	1	0	Hydrazine Addition Pkg.	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.00	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps
D-P100-A803	M-832	1	0	Ammonia Addition Pkg	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.00	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps
D-P100-A803	M-834	1	0	Phosphate Addition Pkg.	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.00	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps
D-P100-A901	M-902	1	0	Cooling Tower System	QCWCAPIT	41,100,000	12,955,985	0.32	$1,659,000	1998	$1,659,000	0.78	$674,181	1.20	$818,659	$682,216	40,000 gpm, 185.4MM BTU/hr
D-P100-A901	M-904	1	0	Plant Air Compressor	STRM0101	159,950	159,950	1.00	$60,100	1997	$60,100	0.34	$60,100	1.30	$79,675	$61,288	450 cfm, 125 psig outlet
D-P100-A901	M-908	1	0	Chilled Water Package	QCHLWCAP	5,040,000	2,268,000	0.45	$380,000	1997	$380,000	0.8	$200,610	1.20	$245,492	$204,577	1000 ton, 600kW
D-P100-A903	M-910	1	0	CIP System	STRM0914	63	28	0.45	$95,000	1995	$95,000	0.6	$58,837	1.20	$73,021	$60,851	designed by Delta-T, (est 0.2 kW)
		38	**2**	**40**										**1.15**	**$ 31,326,762**	**$ 783,169**	
		sum	sum	total										avg.	sum	avg. (installed)	

D	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
D-P100-A201	P-201	1	1	Sulfuric Acid Pump	STRM0710	1,647	414	0.25	$4,800	1997	$9,600	0.79	$3,228	2.79	$9,190	$3,291	2 gpm, 245 ft. head
D-P100-A203	P-209	1	1	Overlimed Hydrolyzate Pump	STRM0228	167,050	102,608	0.61	$10,700	1997	$21,400	0.79	$14,561	2.79	$41,458	$14,849	448 gpm, 150 ft. head
D-P100-A203	P-222	1	1	Filtered Hydrolyzate Pump	STRM0230	162,090	101,614	0.63	$10,800	1997	$21,600	0.79	$14,936	2.79	$42,526	$15,231	448 gpm, 150 ft head
D-P100-A203	P-223	1	0	Lime Unloading Blower	STRM0227	547	337	0.62	$47,600	1998	$47,600	0.5	$37,340	1.40	$52,898	$37,785	3341 cfm, 6 psi, 10,024 lb/hr
D-P100-A202	P-224	1	1	Hydrolysis Feed Pump	STRM0250	160,000	167,795	1.05	$64,934	1999	$129,868	0.6	$133,628	1.20	$160,354	$133,628	740 gpm, 240 ft head
D-P100-A202	P-225	1	1	ISEP Elution Pump	STRM0243	52,731	18,005	0.34	$7,900	1997	$15,800	0.79	$6,761	2.79	$19,249	$6,894	104 gpm, 150 ft head
D-P100-A202	P-226	1	1	ISEP Reload Pump	STRM0246	164,080	100,802	0.61	$8,700	1997	$17,400	0.79	$11,841	2.79	$33,714	$12,075	445 gpm, 150 ft head
D-P100-A202	P-227	1	1	ISEP Hydrolyzate Feed Pump	STRM0221	160,290	98,157	0.61	$10,700	1997	$21,400	0.79	$14,526	2.79	$41,359	$14,814	432 gpm, 150 ft head
D-P100-A203	P-239	1	1	Reacidified Liquor Pump	STRM0239	167,280	102,752	0.61	$10,800	1997	$21,600	0.79	$14,698	2.79	$41,847	$14,988	450 gpm, 100 ft head
D-P100-A302	P-300	4	1	Fermentation Recirc./Transfer Pump	QHX300EA	67,737	55,505	0.82	$8,000	1997	$40,000	0.79	$34,177	2.79	$97,307	$34,852	844 gpm @ 150 ft sized based on heating rate
D-P100-A301	P-301	1	1	Fermentation Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$22,194	1998	$44,388	0.7	$24,168	1.40	$34,238	$24,456	280 gpm @ 150 ft head
D-P100-A301	P-302	2	0	Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$54,088	1998	$108,176	0.7	$58,898	1.40	$83,440	$59,600	504 gpm total, 252 gpm each, 100 ft head
D-P100-A302	P-306	1	1	Beer Transfer Pump	STRM0502	381,701	173,737	0.46	$17,300	1997	$34,600	0.79	$18,579	2.79	$52,899	$18,947	790 gpm each, 171 ft head
D-P100-A307	P-308	8	1	Hydrolyzer Bottoms Pump	STRM0302B	157,136	157,136	1.00	$121,690	1999	$1,095,210	0.6	$1,095,210	1.20	$1,314,252	$1,095,210	3000 GPM each Disc flow pumps, 245ft head
D-P100-A402	P-400	1	1	Cellulase Transfer Pump	STRM0420	40,543	11,600	0.29	$9,300	1997	$18,600	0.79	$6,921	2.79	$19,706	$7,058	58 GPM / 100 ft. head
D-P100-A401	P-401	1	1	Cellulase Seed Pump	STRM0433	2,790	932	0.33	$12,105	1998	$24,210	0.7	$11,236	1.20	$13,644	$11,370	24 gpm / 1 hp
D-P100-A402	P-405	1	1	Media Pump	STRM0416	586	200	0.34	$8,300	1997	$16,600	0.79	$7,104	2.79	$20,227	$7,245	21 Gpm/100 Ft Head
D-P100-A405	P-420	1	1	Anti-foam Pump	STRM0417	227	79	0.35	$5,500	1997	$11,000	0.79	$4,761	2.79	$13,555	$4,855	4 gpm / 75 ft head
D-P100-A501	P-501	1	1	Beer Column Bottoms Pump	P501FLOW	5,053	2,200	0.44	$42,300	1997	$84,600	0.79	$43,861	2.79	$124,881	$44,728	2200 gpm, 150 ft head
D-P100-A501	P-503	1	1	Beer Column Reflux Pump	QCND0501	277,820	131,557	0.47	$1,357	1998	$2,714	0.79	$1,504	2.79	$4,248	$1,522	6 gpm, 140 ft head
D-P100-A502	P-504	1	1	Rectification Column Bottoms Pump	STRM0516	31,507	15,530	0.49	$4,916	1998	$9,832	0.79	$5,622	2.79	$15,884	$5,689	76 gpm, 158 ft head
D-P100-A502	P-505	1	1	Rectification Column Reflux Pump	QCND0502	4,906,301	2,323,304	0.47	$4,782	1998	$9,564	0.79	$5,299	2.79	$14,970	$5,362	207 gpm, 110 ft head
D-P100-A504	P-511	2	1	1st Effect Pump	STRM0525	278,645	133,617	0.48	$19,700	1997	$59,100	0.79	$33,069	2.79	$94,155	$33,723	1137 gpm each, 110 ft head
D-P100-A504	P-512	1	1	2nd Effect Pump	STRM0528	91,111	45,390	0.50	$13,900	1997	$27,800	0.79	$16,032	2.79	$45,646	$16,349	599 gpm, 110 ft head
D-P100-A504	P-513	2	1	3rd Effect Pump	STRM0531	48,001	23,814	0.50	$8,000	1997	$24,000	0.79	$13,795	2.79	$39,276	$14,068	196 gpm each, 110 ft head
D-P100-A504	P-514	1	1	Evaporator Condensate Pump	STRM534A	140,220	69,285	0.49	$12,300	1997	$24,600	0.79	$14,095	2.79	$40,131	$14,374	293 gpm, 125 ft head
D-P100-A502	P-515	1	1	Scrubber Bottoms Pump	STRM0551	15,377	7,427	0.48	$2,793	1998	$5,586	0.79	$3,143	2.79	$8,881	$3,181	31 gpm, 104 ft head
D-P100-A501	P-517	1	1	Kill Tank Bottoms Pump	STRM0518	5,053	660	0.13	$42,300	1997	$84,600	0.79	$16,944	2.79	$48,242	$17,279	660gpm, 72 ft head
D-P100-A601	P-630	1	1	Recycle Water Pump	STRM0602	179,446	84,120	0.47	$10,600	1997	$21,200	0.79	$11,652	2.79	$33,175	$11,882	370 gpm, 150ft head
D-P100-A701	P-703	1	1	Sulfuric Acid Pump	STRM0710	1,647	1,912	1.16	$8,000	1997	$16,000	0.79	$18,001	2.79	$51,253	$18,357	215 gpm, 150ft head
D-P100-A701	P-707	1	1	Antifoam Store Pump	STRM0417	227	79	0.35	$5,700	1997	$11,400	0.79	$4,934	2.79	$14,048	$5,031	0.5 gpm, 92 ft head
D-P100-A701	P-720	1	1	CSL Pump	STRM0735	2,039	859	0.42	$8,800	1997	$17,600	0.79	$8,889	2.79	$25,308	$9,065	182 gpm, 150ft head
D-P100-A802	P-804	2	1	Condensate Pump	STRM811A	249,633	38,798	0.16	$7,100	1997	$21,300	0.79	$4,894	4.60	$22,958	$4,991	130 gpm, 150' head
D-P100-A802	P-824	2	1	Deaerator Feed Pump	STRM811A	196,000	38,798	0.20	$9,500	1997	$28,500	0.79	$7,927	8.30	$67,097	$8,084	180 gpm, 115' head
D-P100-A802	P-826	4	1	BFW Pump	STRM0813	207,310	80,536	0.39	$52,501	1998	$262,505	0.79	$124,377	1.40	$176,203	$125,859	310 gpm, 2740' head
D-P100-A802	P-828	1	1	Blowdown Pump	STRM0821	6,600	2,699	0.41	$5,100	1997	$10,200	0.79	$5,032	6.40	$32,842	$5,132	12 gpm, 150' head
D-P100-A803	P-830	1	1	Hydrazine Transfer Pump	STRM813A	229,386	80,536	0.35	$5,500	1997	$11,000	0.79	$4,811	6.40	$31,402	$4,907	3 gpm, 75' head
D-P100-A901	P-902	1	1	Cooling Water Pumps	STRM0940	18,290,000	5,553,791	0.30	$332,300	1997	$664,600	0.79	$259,201	2.79	$737,993	$264,326	12300 gpm, 70ft head
D-P100-A902	P-912	1	1	Make-up Water Pump	STRM0904	244,160	82,445	0.34	$10,800	1997	$21,600	0.79	$9,161	2.79	$26,084	$9,343	370 gpm, 75ft head
D-P100-A902	P-914	1	1	Process Water Circulating Pump	STRM0905	352,710	111,503	0.32	$11,100	1997	$22,200	0.79	$8,938	2.79	$25,449	$9,115	745 gpm, 75ft head
		58	**38**	**96**										**2.90**	**$ 3,771,987**	**$ 39,292**	
		sum	sum	total										avg.	sum	avg. (installed)	

D	Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Instal. Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description
D-P100-A202	S-202	3	0	Pre-IX Belt Filter Press	SOLD0220	57,000	57,000	1.00	$200,000	1998	$600,000	0.39	$600,000	1.40	$850,010	$607,150	Use 3 units for 45% of the flow as recommended by the vendor
D-P100-A202	S-221	1	0	ISEP	STRM0240	210,005	98,157	0.47	$2,058,000	1997	$2,058,000	0.33	$1,601,194	1.20	$1,959,422	$1,632,851	10 chambers (39" dia. X 84" high), 4" dia. Valve - Weak Base Resin
D-P100-A203	S-222	1	0	Hydroclone & Rotary Drum Filter	STRM0229	5,195	1,137	0.22	$165,000	1998	$165,000	0.39	$91,224	1.40	$129,235	$92,311	Hydrocyclone and Vacuum Filter for 453 gpm
D-P100-A203	S-227	1	0	LimeDust Vent Baghouse	STRM0227	548	337	0.61	$32,200	1997	$32,200	1	$19,778	1.50	$30,254	$20,169	3750 cfm, 625 sf, 6 cfm/sf
D-P100-A601	S-601	2	0	Beer Column Bottoms Centrifuge	CENTFLOW	404	300	0.74	$659,550	1998	$1,319,100	0.6	$1,103,371	1.20	$1,339,824	$1,116,520	requires 540gpm duty, 2 @ 300 gpm and 410 hp each
D-P100-A901	S-904	1	1	Instrument Air Dryer	STRM0101	159,950	71,977	0.45	$15,498	1999	$30,996	0.6	$19,197	1.30	$24,956	$19,197	134 scfm air dryer, -40F Dewpoint
		9	1	10										1.33	$ 4,333,701	$ 433,370	
		sum	sum	total										avg.	sum	avg. (installed)	
D-P100-A201	T-201	1	0	Sulfuric Acid Storage	STRM0710	1,647	860	0.52	$5,760	1996	$5,760	0.71	$3,633	1.68	$6,283	$3,751	2000 gal., 24 hr. residence time, 90% wv, 5.5ft diam. X 11ft
D-P100-A201	T-203	1	0	Blowdown Tank	STRM0217	270,300	121,514	0.45	$64,100	1997	$64,100	0.93	$30,475	1.68	$52,061	$31,078	7000 gal., 11' dia x 30' high, 10 min. res. time, 75% wv, 15 psig
D-P100-A203	T-209	1	0	Overliming Tank	STRM0228	167,050	102,608	0.61	$71,000	1997	$71,000	0.71	$50,232	1.76	$90,186	$51,225	29850 gal., 16' dia. X 32' high, 1 hr. res. time, 90% wv, 15 psig
D-P100-A203	T-220	1	0	Lime Storage Bin	STRM0227	548	548	1.00	$69,200	1997	$69,200	0.46	$69,200	1.76	$124,243	$70,568	4455 cf, 14' dia x 25' high, 1.5x rail car vol., atmospheric, 15 day storage max
D-P100-A203	T-224	1	0	Reacidification Tank	STRM0239	102,752	102,752	1.00	$111,889	1999	$111,889	0.51	$111,889	1.76	$196,992	$111,889	120,000 gal., 28' dia x 28' high, 4 hr. res. time, 90% wv, atmospheric
D-P100-A202	T-232	1	0	Slurrying Tank	STRM0250	358,810	167,795	0.47	$44,800	1997	$44,800	0.71	$26,117	1.76	$46,890	$26,633	11300 gal., 13' dia. X 25' high, 15 min. res. time, 90% wv
D-P100-A301	T-301	1	0	Fermentation Seed Hold Tank	STRM0304	41,777	17,529	0.42	$161,593	1998	$161,593	0.51	$103,767	1.76	$184,870	$105,003	105000 gal., API atmospheric
D-P100-A302	T-306	1	0	Beer Well	STRM0502	129,000	183,467	1.42	$111,889	1999	$111,889	0.51	$133,906	1.76	$235,756	$133,906	192,518 gal., 32' dia x 32' high, 4 hr. res. time, 95% wv, atmospheric
D-P100-A307	T-307	4	0	Enzymatic Hydrolysis Tank	STRM0302B	750,000	375,000	0.50	$326,203	1999	$1,304,812	0.6	$860,855	2.04	$1,753,728	$860,855	375,000 gallons, 24 hour residence time, 2 side mounted agitators cone bottom, concrete base, bottom outlet through the concrete, 30o cone bottom
D-P100-A302	T-405	1	0	Media-Prep Tank	STRM0416	586	200	0.34	$64,600	1997	$64,600	0.71	$30,128	1.68	$51,467	$30,723	2083 Gal / 1.17 hp Agitator
D-P100-A402	T-420	1	0	Anti-foam Tank	STRM0417	227	79	0.35	$402	1998	$402	0.71	$189	1.68	$321	$192	67 gal, 3 hr. residence time
D-P100-A501	T-503	1	0	Beer Column Relfux Drum	QCND0501	277,820	131,557	0.47	$11,900	1997	$11,900	0.93	$5,938	1.68	$10,144	$6,055	164 gal, 15 min res. Time, 50% wv, 2'6" dia., 5' long, 25 psig
D-P100-A502	T-505	1	0	Rectification Column Reflux Drum	QCND0502	4,906,301	2,323,304	0.47	$45,600	1997	$45,600	0.72	$26,621	1.68	$45,476	$27,147	6225 gal, 15 min res time, 50% wv, 7' dia, 22' long, 25 psig
D-P100-A502	T-512	1	0	Vent Scrubber	STRM0523	18,523	9,788	0.53	$99,000	1998	$99,000	0.78	$60,197	1.68	$102,043	$60,915	5' dia x 25' high, 4 stages, plastic Jaeger Tri-Packing
D-P100-A501	T-513	1	0	Kill Tank	STRM0518	149,897	149,897	1.00	$99,920	1999	$99,920	0.78	$99,920	1.68	$167,384	$99,920	18 psig, 30 min. res. time
D-P100-A601	T-630	1	0	Recycled Water Tank	STRM0602	179,446	84,120	0.47	$14,515	1998	$14,515	0.745	$8,254	1.68	$13,992	$8,353	7410 gal, 20 min. res., 2.5 psig, 9.5ft diam. x 14.25ft
D-P100-A701	T-703	1	0	Sulfuric Acid Storage Tank	STRM0710	1,647	1,912	1.16	$42,500	1997	$42,500	0.51	$45,860	1.76	$82,338	$46,767	20,000 gal, 240 hr supply, 90% wv, 12ft diam. x 24 ft, atmospheric
D-P100-A701	T-707	1	0	Antifoam Storage Tank	STRM0417	227	227	1.00	$14,400	1997	$14,400	0.71	$14,400	1.68	$24,600	$14,685	12,000 gal, 27 day supply, 10.5ft diam. X 18.5ft
D-P100-A701	T-720	1	0	CSL Storage Tank	STRM0735	2,039	859	0.42	$88,100	1997	$88,100	0.79	$44,495	1.68	$76,011	$45,375	30160 gal, 90% wv, 120 supply, 14.3ft diam. X 25 ft
D-P100-A802	T-804	1	0	Condensate Collection Tank	STRM811A	229,386	38,798	0.17	$7,100	1997	$7,100	0.71	$2,011	3.30	$6,766	$2,050	200 gal, 1.5 min. res. time
D-P100-A802	T-824	1	0	Condensate Surge Drum	STRM811A	150,000	38,798	0.26	$49,600	1997	$49,600	0.72	$18,734	5.00	$95,523	$19,105	2100 gal., 6' diam. X 10', 15 psig, res. time 11 min.
D-P100-A802	T-826	1	0	Deaerator	STRM0813	267,000	80,536	0.30	$165,000	1998	$165,000	0.72	$69,616	6.50	$457,896	$70,446	3030 gal., 15 psig, 10 min. res.
D-P100-A802	T-828	1	0	Blowdown Flash Drum	STRM0821	6,550	2,699	0.41	$9,200	1997	$9,200	0.72	$4,859	7.30	$36,168	$4,955	210 gal., 2.5' diam. X 6', 50 psig 17 min. res.
D-P100-A803	T-830	1	0	Hydrazine Drum	STRM813A	229,386	80,536	0.35	$12,400	1997	$12,400	0.93	$4,685	7.00	$33,440	$4,777	138 gal, 3.75' x 1.25' diam., 10 psig
D-P100-A901	T-904	1	0	Plant Air Receiver	STRM0101	159,950	53,316	0.33	$13,000	1997	$13,000	0.72	$5,894	1.68	$10,069	$6,011	300 gal., 200 psig
D-P100-A902	T-914	1	0	Process Water Tank	STRM0905	352,710	111,503	0.32	$195,500	1997	$195,500	0.51	$108,663	1.76	$195,095	$110,811	234360 gal, 8hr res. time
		29	0	29										2.51	$ 4,099,742	$ 141,370	
		sum	sum	total										avg.	sum	avg. (installed)	

TOTAL TAG ITEMS:	155
TOTAL PIECES:	310

Appendix 6

PHOENIX BIO-SYSTEMS, INC.

at ICM, Inc.:

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Colwich, Kansas 67030
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NREL Corn Stover to Ethanol – High Plains Fuel Ethanol Addition – Wastewater – Revision 1

Wastewater Analysis – 98,000 kg/hr total Flow

The attached mass balance estimate describes the proposed wastewater from corn stover processing at the High Plains Plant in York, NE. The overall wastewater, originally described by Merrick Engineering, has been further divided into components as they relate to biological digestion in an anaerobic wastewater treatment system. The stream given by Merrick is the sum of streams 520-Flash to WW Treatment, 247-IX to WW Treatment, 535- to WW Treatment and feedstock receiving pad run-off.

Fate of Components

There are several areas worthy of consideration in the analysis. The main organic components of this stream are ammonium acetate, acetic acid, ethanol, furfural and HMF. It has been assumed for the purposes of this analysis that these are all amenable to anaerobic digestion to some extent. The acetate and ethanol components are assumed to be 98 percent removable, while furfural and HMF are assumed to be 80 percent removable. The Corn Steep Liquor is assumed to be 92 percent removable as well.

Aerobic removals of residuals after anaerobic digestion are considered to be better, averaging 98%. Anaerobic digestion is chosen as the least cost method for removal of the largest components of organic COD.

All of the organic components have been expressed as their equivalent Chemical Oxygen Demand (COD) for complete conversion to carbon dioxide and water. Furthermore, values have been converted to pounds per day, which gives the average American reader a better "feel" for the amounts derived. It can be seen that the organic components alone generate approximately 72,000 pounds per day of COD for anaerobic digestion,

This amount of COD also generates some 437,000 cubic feet per day of biogas.

Sulfate

There is a very significant amount of sulfate included in this stream, due primarily to the need for sulfuric acid regeneration of IX resin used for the removal of Acetic Acid from the Hydrolysis-Fermentation stream. If all of the sulfate were to be converted to hydrogen sulfide in anaerobic digestion, then some 5,699 pounds per day of hydrogen sulfide would be produced. That is the equivalent of 138,000 ppm v/v in the biogas. However, in practice, anaerobic digesters fed very high sulfate streams appear to be self-limiting in hydrogen sulfide production. Hydrogen sulfide in biogas rarely exceeds 5,000 ppm v/v. For the purposes of this analysis, it was assumed that no more than 5,000 ppm v/v H2S would actually occur in the biogas. Therefore only a small percentage of the available sulfate from ammonium sulfate was theoretically converted to H2S. The remainder is carried through the process as the salt of ammonia.

Ammonia Nitrogen

Ammonia is also very high in this waste stream, also due to the IX process regeneration. The hydrolysis of ammonium acetate in the digester results in over 9,300 pounds per day of Ammonia-Nitrogen which, when considered as COD, demands over 40,000 pounds per day of oxygen for conversion to nitrate. Anaerobic digestion will not remove this ammonia nitrogen but will pass it through the reactor in solution.

Among the options for treating this residual ammonia are air stripping and nitrification. Air stripping may be accomplished either during anaerobic digestion or afterward. It should be noted that 9,300 pounds per day of ammonia is likely to be a significant source of air emissions (4.6 tons per day of ammonia is equal to over 1,500 tons per year).

Nitrification is likely to be a more practical treatment, however, it will require some 40,000 pounds per day of oxygen for conversion to nitrate.

Secondary Treatment

Secondary aerobic treatment will be required in order to address both the residual ammonia and some 8,000 pounds per day of residual organics from the anaerobic digester.

Existing Capacity at High Plains – York

The existing waste water treatment plant at the High Plains Plant consists of a Bio-Methanation Anaerobic digester, a 2.6 million gallon aerobic lagoon with return activated sludge capability and 400 horsepower aeration, a sludge clarifier, and a sludge holding and aeration pond.

Wastewater is currently pre-treated in this system for discharge to the City of York. It is expected that the City of York might be capable of managing the hydraulic load from the corn stover process, but will impose stringent limits on COD, TSS, and Ammonia Nitrogen.

The existing anaerobic system at York is capable of 18,000 pounds per day of COD removal. Currently, the plant at York utilizes 50 to 75% of this capacity. Therefore, it will be necessary to add significant anaerobic pre-treatment for the corn stover process. Approximately 500,000 gallons of anaerobic digestion capacity will be required.

Although the existing aerobic system may be capable of treating a portion of the anaerobic effluent from the corn stover waste water, significant additional aerobic capacity will be required. The equivalent of at least 40,000 pounds per day of COD removal would be prudent. Furthermore, clarification and sludge management facilities would also require expansion.

Estimated Expansion Requirements

An equalization basin will be required with capacity no less than 300,000 gallons. An above-ground bolted tank with a cover, including foundations, pumps and controls is estimated to cost **$0.35 million**. The equalization basin is sized to accommodate approximately one half day flow. Flow would proceed from equalization to the anaerobic system.

Anaerobic digestion will require 500,000 gallons of additional capacity. Estimated cost of expansion is **$3.2 million**, including site work, foundations, reactors and ancillary equipment.

Expansion of aerobic facilities can be accomplished with the addition of four 350,000 gallon Sequencing Batch Reactors, with a capacity of 48,000 pounds per day of oxygen transfer, along with de-nitrification capability. Aeration and mixing would require approximately 1,400 horsepower. Estimated cost for the aerobic section of the expanded plant is **$4.3 million.**

Expansion of clarification facilities would not be required as Sequencing Batch Reactors also act as clarifiers during the "Settling Phase".

The City of York is unlikely to accommodate wastewater with nitrate concentrations approximating 2,870 mg/l, therefore, de-nitrification capability would be required. Residual ammonia totals over 9,570 ppd or 835 mg/l, and when converted to nitrate, will be over 2,870 mg/l (32,920 ppd). Conventional means of de-nitrification, such as single or double sludge de-nitrification are likely not adequate for this task, however, Sequencing Batch Reactors have inherent de-nitrification capability. Inclusion of an "anoxic" phase in the Batch sequence converts nitrate to nitrogen gas.

Final filtration through pressure sand filters is recommended. Pressure Sand Filters with 200 square feet of filtration surface area would suffice. This system would consist of 4 x 8'd pressure sand filters, stainless steel construction, with auto-backwash, in a small building. The estimated cost for this system is **$0.28 million.**

Summarizing capital costs;

- Equalization, one 300,000 gal eq. Tank- **$ 0.35 M**
- Anaerobic System, as above- **$3.2 M**
- Aerobic SBR's – 4 x 350,000 gal - **$4.3 M**
- Filters- 200 sq. ft. - **$0.28 M**
-

Total cost of capital improvements without NPDES discharge is estimated to be **$8.13 million.**

The current PFD for the corn stover operation calls for the recycling of wastewater to process use. This is feasible provided that there is sufficient water removed from the process to provide adequate desalting of the total process water. Final sand filtration is recommended for this case. There will be approximately 4,400 mg/l of inorganic salts in the recycle water. This concentration, approaching 0.5% brine could be problematic for re-use. With 50% dilution from fresh water the risk of salting the process is reduced considerably.

In the event that the wastewater cannot be re-used, the city of York may not accommodate the hydraulic flow (622,000 gpd) created by the corn stover process. Current hydraulic flow from existing facility averages over 350,000 gpd, including cooling tower blow-down. An NPDES permit may be required for direct discharge of the additional wastewater.

With NPDES discharge of wastewater capital cost is likely to rise for the cost of out-fall, monitoring stations and additional engineering/legal expenses. Operating costs would also increase due to increased monitoring.

ADDENDUM (10-19-99)

Reduced Hydraulic Flow

Closer review of the various streams comprising the wastewater stream for this project indicates that there will be significantly less wastewater volume than originally believed. Current mass balance for the processing facility indicates an average flow of 98,267 kg/hr versus the original flow of 217,300 kg/hr. The difference is apparently due to an overestimate of run-off from the feedstock delivery pad during storm events. Correction of this estimate and leveling of storm water flow to the wastewater treatment system results in a much lower total flow.

Unfortunately, the reduced hydraulic load has little impact on the size requirements of both the anaerobic and aerobic treatment units. The reason for this is that all of the organic and nitrogenous wastes are carried by the other plant streams.

The equalization basin and the aerobic SBR system can be reduced in size in accord with the lower hydraulic flow. The information given above is valid for the reduced flow case.

Removal of IX Treatment

It has been suggested that the Ion Exchange removal of Acetic acid might be eliminated from the proposed process. If research shows this to be possible, the savings in wastewater treatment and chemical costs would be significant.

Although Acetic Acid would still be produced in Stover hydrolysis, if it could be successfully carried through fermentation, it would be removed in the anaerobic reactor. This water could be recycled without the risk of acetate poisoning of the yeast fermentation.

Furthermore, the deletion of IX would eliminate the requirement for the purchase and application of Ammonia for regeneration. This would also remove the requirement for over 40,000 ppd of oxygenation for nitrogen removal in the aerobic wastewater treatment system.

Mass Balance and operating cost estimates have been completed for both of these cases;
1- Reduced hydraulic flow to 98,000 kg/hr and
2- Elimination of IX treatment for acetic acid removal.

It is obvious that the elimination of IX treatment has very significant economic impact on operating costs for wastewater treatment. The net operating cost of treatment for the reduced flow case (including credit for biogas produced) is **$913,000** per year without depreciation. The net operating cost for treatment without IX is **$122,000** per year. Net savings is $791,000 per year or 87% of operating costs. The difference is due to reduced operating costs associated with the removal of Ammonia-derived nitrogen from the wastewater.

In addition, capital costs will be lower due to the need for much less aerobic capacity. The aerobic section of wastewater treatment can be reduced from 4 x 350,000 gal SBR's to 2 x 180,000 gal SBR's. Aeration systems will be reduced as well. Capital cost for the reduced aerobic SBR system is estimated at **$1.73 M.** Capital for all components of this system would be;

- Equalization, one 200,000 gal eq. Tank- **$ 0.295 M**
- Anaerobic System, as above- **$3.2 M**
- Aerobic SBR's **- $1.23 M**
- Filters- 150 sq. ft. - **$0.245 M**

The total capital for this complete system would be **$4.97 M,** which is a capital savings of 39%.

Some Caustic has been included in the operating costs for this case since ammonia nitrogen is no longer in high concentration. In the earlier case no caustic was required due to the presence of large amounts of ammonia.

In addition, sulfate is no longer a problem as the elimination of IX and associated sulfuric acid has reduced available sulfate to what would be derived from feedstock and make-up fresh water. It is expected that Hydrogen sulfide would not exceed 500 ppm in the biogas, which is easily removed with low cost scrubbing.

Salt concentration in this treated wastewater would be quite low and would pose no significant risk for re-use.

This system would be capable of ;

a- Producing water for discharge to the environment

b- Producing water for discharge to the City of York without surcharge

c- Producing water for re-use in the process

Cooling Tower Blow-Down

Cooling tower blow-downs have been deleted from this analysis since these waters do not contain appreciable amounts of pollutants. Generally, cooling tower blow-downs can be released to the environment on NPDES permits, without difficulty.

Operating Costs - NREL - Corn Stover Wastewater Model - 98,000 kg/hr to WW - WITHOUT IX

PARAMETERS	ANAEROBIC BIO-METHANATOR AMOUNTS	ANAEROBIC BIO-METHANATOR DAILY COST	DISCHARGE WITH SBR AEROBIC TREATMENT AMOUNTS	DISCHARGE WITH SBR AEROBIC TREATMENT DAILY COST
Flow, Gallons Per Minute (GPM)	301.00		301.00	
Flow, Gallons Per Day (GPD)	433,440.00		433,440.00	
Chemical Oxygen Demand (COD) mg/l	20,000.00		1,800.00	
Biological Oxygen Demand (BOD5) mg/l	12,000.00		1,080.00	
Pounds Per Day COD	72,271.82		6,504.46	
Pounds Per Day BOD	50,590.28		4,553.12	
Inlet Temperature	30C		30C	
Total Nitrogen mg/l	250.00		205.00	
Total Nitrogen PPD	903.40		740.79	
Total Phosphate mg/l	30.00		28.00	
Total PhosphatePPD	108.41		101.18	
COD Space Loading Rate g/l/d	18.00		2.00	
COD Reduction	0.93		0.98	
Residual COD mg/l	1,400.00		36.00	
Residual COD PPD	5,059.03		130.09	
Residual BOD5 mg/l	840.00		10.80	
Residual BOD5 PPD	3,035.42		91.06	
TSS mg/l	0.00		100.00	
TSS PPD	0.00		361.36	
Horsepower Required:				
Blower Horsepower	5.00		150.57	
Mixing	0.00		67.75	
Pumping	34.68		43.34	
Total Horsepower	39.68		261.67	
Cost per kwh	0.035		0.035	
Kwh per day	704.63	$24.66	4,647.17	$162.65
Chemicals Required, lbs/day:				
Nitrogen	(773.31)	$0.00	0.00	$0.00
Phosphate	(65.04)	$0.00	0.00	$0.00
Micro-Nutrients	7.23	$3.61	0.00	$0.00
Caustic lbs/day Required	328.11	$49.22	0.00	$0.00
Polymer @ $ 2.50/lb	0.00	$0.00	35.00	$87.50
Chlorine	0.00	$0.00	0.00	$0.00
Sludge (Biomass) Generation:				
Dry Weight Yield, lbs/day	1,445.44		1,951.34	
Wet Weight of Sludge, lbs/day	24,090.61		195,133.92	
Sludge Total Solids	6%		1%	
Sludge Yield on COD	2%		30%	

Operating Costs - NREL - Corn Stover Wastewater Model - 98,000 kg/hr to WW - WITHOUT IX

PARAMETERS	AMOUNTS	DAILY COST	AMOUNTS	DAILY COST
Sludge Disposal :				
Dewatering @ $ 0.XX per 1000 lb wet weight	0.00	$0.00	0.00	$0.00
Volume Reduction	0%		0.00	$0.00
Disposal Volume-gal	0.00		23,369.33	
Disposal @ $ 0.0X/gal	0.00	$0.00	0.010	$233.69
Bio-Gas Produced (CFD):	443,987.25		0.00	
Methane Yield (85%) CFD	377,389.16		0.00	
Less Heating Requirement	0.00		0.00	
Net Methane for energy- CFD	377,389.16		0.00	
Bio-Gas Credit ($2.50/MMBTU Methane)		($943.47)	0.00	$0.00
Labor:				
Cost per hour ($)	18.00		18.00	
Manhours / Day	3.00	$54.00	8.00	$144.00
Maintenance parts	50.00	$60.00	50.00	$60.00
Sewer Surcharge (if applicable):				
Flow @ $0.XX /1000 gal	0.00	$0.00	1.00	$433.44
Allowable BOD5 Concentration mg/l	300.00		300.00	
PPD Allowable BOD5	1,084.08		1,084.08	
Residual BOD5 to Sewer PPD	1,951.34		(993.01)	
BOD5 Surcharge @ $x.xx/lb	0.20	$0.00	0.20	$0.00
Allowable TSS Concentration mg/l	250.00		250.00	
PPD Allowable TSS	903.40		903.40	
Residual TSS to Sewer PPD	0.00		361.36	
TSS @ $0.XX /lb	0.00	$0.00	0.00	$0.00
Total Daily Cost		**($751.98)**		**$1,121.28**
Annual Cost (Days per year)	330.00	**($248,153.43)**	330.00	**$370,023.84**
Daily Operating Cost w/o Methane Credit		**$191.49**		**$1,121.28**
Annual Operating Cost w/o Methane Credit	330.00	**$63,192.63**	330.00	**$433,216.47**

Operating Costs - NREL - Corn Stover Wastewater Model - 98,000 kg/hr to WW

	ANAEROBIC BIO-METHANATOR		DISCHARGE WITH SBR AEROBIC TREATMENT	
PARAMETERS	AMOUNTS	DAILY COST	AMOUNTS	DAILY COST
Flow, Gallons Per Minute (GPM)	432.00		432.00	
Flow, Gallons Per Day (GPD)	622,080.00		622,080.00	
Chemical Oxygen Demand (COD) mg/l	13,700.00		9,450.00	
Biological Oxygen Demand (BOD5) mg/l	8,220.00		5,670.00	
Pounds Per Day COD	71,052.09		49,010.38	
Pounds Per Day BOD	49,736.46		34,307.27	
Inlet Temperature	30C		30C	
Total Nitrogen mg/l	2,959.00		2,920.00	
Total Nitrogen PPD	15,346.21		15,143.95	
Total Phosphate mg/l	30.00		28.00	
Total PhosphatePPD	155.59		145.22	
COD Space Loading Rate g/l/d	18.00		4.00	
COD Reduction	0.93		0.98	
Residual COD mg/l	959.00		189.00	
Residual COD PPD	4,973.65		980.21	
Residual BOD5 mg/l	575.40		56.70	
Residual BOD5 PPD	2,984.19		686.15	
TSS mg/l	0.00		100.00	
TSS PPD	0.00		518.63	
Horsepower Required:				
Blower Horsepower	5.00		1,134.50	
Mixing	0.00		204.21	
Pumping	49.77		62.21	
Total Horsepower	54.77		1,400.92	
Cost per kwh	0.035		0.035	
Kwh per day	972.65	$34.04	24,880.29	$870.81
Chemicals Required, lbs/day:				
Nitrogen	(15,218.32)	$0.00	0.00	$0.00
Phosphate	(112.96)	$0.00	0.00	$0.00
Micro-Nutrients	7.11	$3.55	0.00	$0.00
Caustic lbs/day Required	0.00	$0.00	0.00	$0.00
Polymer @ $ 2.50/lb	0.00	$0.00	35.00	$87.50
Chlorine	0.00	$0.00	0.00	$0.00
Sludge (Biomass) Generation:				
Dry Weight Yield, lbs/day	1,421.04		14,703.11	
Wet Weight of Sludge, lbs/day	23,684.03		1,470,311.43	
Sludge Total Solids	6%		1%	
Sludge Yield on COD	2%		30%	

Operating Costs - NREL - Corn Stover Wastewater Model - 98,000 kg/hr to WW

PARAMETERS	AMOUNTS	DAILY COST	AMOUNTS	DAILY COST
Sludge Disposal :				
Dewatering @ $ 0.XX per 1000 lb wet weight	0.00	$0.00	0.00	$0.00
Volume Reduction	0%		0.00	$0.00
Disposal Volume-gal	0.00		176,085.20	
Disposal @ $ 0.0X/gal	0.00	$0.00	0.010	$1,760.85
Bio-Gas Produced (CFD):	436,494.04		0.00	
Methane Yield (85%) CFD	371,019.94		0.00	
Less Heating Requirement	0.00		0.00	
Net Methane for energy- CFD	371,019.94		0.00	
Bio-Gas Credit ($2.50/MMBTU Methane)		($927.55)	0.00	$0.00
Labor:				
Cost per hour ($)	18.00		18.00	
Manhours / Day	3.00	$54.00	8.00	$144.00
Maintenance parts	50.00	$60.00	50.00	$60.00
Sewer Surcharge (if applicable):				
Flow @ $0.XX /1000 gal	0.00	$0.00	1.00	$622.08
Allowable BOD5 Concentration mg/l	300.00		300.00	
PPD Allowable BOD5	1,555.89		1,555.89	
Residual BOD5 to Sewer PPD	1,428.30		(869.74)	
BOD5 Surcharge @ $x.xx/lb	0.20	$0.00	0.20	$0.00
Allowable TSS Concentration mg/l	250.00		250.00	
PPD Allowable TSS	1,296.57		1,296.57	
Residual TSS to Sewer PPD	0.00		518.63	
TSS @ $0.XX /lb	0.00	$0.00	0.00	$0.00
Total Daily Cost		**($775.95)**		**$3,545.24**
Annual Cost (Days per year)	330.00	**($256,064.97)**	330.00	**$1,169,929.96**
Daily Operating Cost w/o Methane Credit		**$151.60**		**$3,545.24**
Annual Operating Cost w/o Methane Credit	330.00	**$50,026.48**	330.00	**$1,219,956.45**

NREL Wastewater- Corn Stover Case - Mass Balance Estimate - 98,000 kg/hr Flow - WITHOUT IX

Wastewater Components	Total Wastewater - Kg/hr	Total Wastewater - Lbs/day	Conc - mg/l	Conc as COD - mg/l	COD - Lbs/day	COD for Anaerobic Digestion - Lbs/day	Residual COD - Lbs/day	Other Residuals - Lbs/day	Residuals Conc - as COD - mg/l	After Aerobic & Deni mg/l
Total Flow-	68,635.0	3,623,928.0				3,623,928.0	3,623,928.0			3,623,928.0
Gallons		434,523.7				434,523.7	434,523.7			434,523.7
Insoluble solids (is)	0.0	0.0								
Soluble solids (ss)	1,297.1	68,486.9	18,905.3							
Water	67,337.9	3,555,441.1								
Ethanol	22.0	1,161.6	320.7	352.7	1,277.8	1,277.8	25.6		7.1	0.1
CSL (ss)	33.0	1,742.4	481.0	529.1	1,916.6	1,916.6	153.3		42.3	0.8
(NH4)2SO4 for digestion	0.0	0.0								
SO4 for conversion from Amm sulfate**	0.0	0.0	0.0						0.0	
NH4 from Amm Sulfate		0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0
Unconverted (NH4)2SO4	0.0	0.0	0.0					0.0		0.0
NH4 from Amm Acetate	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0
Total Acetate C2H4O2	693.1	36,595.7	10,102.0	11,112.2	40,255.2	40,255.2	805.1		222.2	4.4
Furfural	457.0	24,129.6	6,660.8	8,659.0	26,542.6	26,542.6	5,308.5		1,465.4	44.0
HMF	31.0	1,636.8	451.8	587.4	1,800.5	1,800.5	360.1		99.4	3.0
NH4	4.0	211.2	58.3	250.7	908.2		908.2		250.7	7.5
NH4OH***	54.0	2,851.2	787.1					333.0	64.0	1.9
Other	3.0	158.4	43.7	48.1	174.2	174.2	17.4		4.8	2.4
TOTALS			18,905.3	21,539.2	**72,875.1**	**71,966.8**	**7,578.2**	333.0	2,155.9	64.2
Hydrogen Sulfide - *								0.0		
BioGas Production- CFD- 85% CH4						442,113.6				
Energy mmbtu/day						375.8				

* H2S at 5,000 v/v ppm in biogas

** SO4 estimated limit of conversion at 5,000 ppm H2S- in biogas

*** Expected to be neutral salts in digester

NREL Wastewater- Corn Stover Case - Mass Balance Estimate - 98,000 kg/hr Flow

Wastewater Components	Total Wastewater - Kg/hr	Total Wastewater - Lbs/day	Conc - mg/l	Conc as COD - mg/l	COD - Lbs/day	COD for Anaerobic Digestion - Lbs/day	Residual COD - Lbs/day	Other Residuals - Lbs/day	Residuals Conc - as COD - mg/l	After Aerobic & Deni mg/l
Total Flow-	98,267.0	5,188,497.6				5,188,497.6	5,188,497.6			5,188,497.6
Gallons		622,122.0				622,122.0	622,122.0			622,122.0
Insoluble solids (is)	0.0	0.0								
Soluble solids (ss)	1,907.0	100,689.6	19,413.3							
Water	96,360.0	5,087,808.0								
Ethanol	22.0	1,161.6	224.0	246.4	1,277.8	1,277.8	25.6		4.9	0.0
CSL (ss)	33.0	1,742.4	335.9	369.5	1,916.6	1,916.6	153.3		29.6	0.6
(NH4)2SO4 for digestion	16.0	844.8								
SO4 for conversion from Amm sulfate**	11.5	607.2	117.1						0.0	
NH4 from Amm Sulfate	4.5	237.6	45.8	197.0	1,021.7		1,021.7	237.6	197.0	5.9
Unconverted (NH4)2SO4	401.0	21,172.8	4,082.2					21,172.8		4,082.2
NH4 from Amm Acetate	176.9	9,340.3	1,800.8	7,743.6	40,163.4		40,163.4	9,340.3	7,743.6	232.3
Total Acetate C2H4O2	693.1	36,595.7	7,055.8	7,761.3	40,255.2	40,255.2	805.1		155.2	3.1
Furfural	457.0	24,129.6	4,652.3	6,047.9	26,542.6	26,542.6	5,308.5		1,023.5	30.7
HMF	31.0	1,636.8	315.6	410.3	1,800.5	1,800.5	360.1		69.4	2.1
NH4	4.0	211.2	40.7	175.1	908.2		908.2		175.1	15.8
NH4OH***	54.0	2,851.2	549.7					549.7	0.0	106.0
Other	3.0	158.4	30.5	33.6	174.2	174.2	17.4		3.4	1.7
TOTALS			19,250.4	22,984.7	**114,060.1**	**71,966.8**	**48,763.2**	31,300.4	9,401.7	4,480.4
Hydrogen Sulfide - *								215.3		
BioGas Production- CFD- 85% CH4						442,113.6				
Energy mmbtu/day						375.8				

* H2S at 5,000 v/v ppm in biogas

** SO4 estimated limit of conversion at 5,000 ppm H2S- in biogas

*** Expected to be neutral salts in digester

Appendix 7

Structure of Appendix 7

Proforma

Sensitivity Analysis

Cellulase Source Study

<u>Comparison of On-site cellulase production methods</u>

<u>Comparison of On-site and Purchased Cellulase</u>

<u>Method A: "BASED ON PUREVISION LABORATORY RESULTS OF COMPARISON"</u>

<u>Method B: "BASED ON PRODUCT SPECIFICATIONS PROVIDED BY SPECIALTY ENZYMES INC."</u>

Proforma

High Plains Corp.
York, NE Co-located
Stover-to-Ethanol Plant

A
1/26/00

EL ENZYMATIC HYDROLYSIS - PRO FORMA

erlying Assumptions & Input Variables

CURRENT SITUATION:

The Pro Forma models an Enzymatic Hydrolysis Ethanol plant using corn stover as the feed stock.

ETHANOL

The plant will convert corn stover to fuel grade ethanol utilizing enzymatic hydrolosis.

Corn stover feed rate of		71,977	kg/hr (str 101), produce estimated total output in				
equivalent kilograms of fuel grade ETOH			9,151	kg/hr.	=	76,871,691	kg / year (str 515)
gal./short ton=	74.1		3,065	gal/hr	=	25,746,124	gal / year
gal./metric ton=	81.7						

Increase to current York yearly production: 70%

The model assumes renewal of the ethanol excise tax credit of $.54 per gallon to the blender and **NOT** the small producer tax credit of $.10 per gallon through the year 2015 for a total ethanol value of

$1.10 per gallon or $0.37 per kg and **$ 28,320,736 per year TOTAL Ethanol sales**

CARBON DIOXIDE

Currently, carbon dioxide from the High Plains York fermentations is sold to a CO_2 compression company.

Diverting the CO_2 (stm 550) from the stover plant into this stream for sale as opposed to the atmosphere provides

110,749 kg/hr = 930,294 ton / year with a value of $ 4.13 per metric ton

WITH THIS PROFORMA NO CO_2 IS SOLD. **CO_2 Value/year = $0**

LIGNIN

A Lignin co-product is produced and sold as combustion fuel material. A total amount of lignin in the stream (stm 601B) is

63,778 kg/hr = 535,734 metric ton / year is produced from the process.

The water in the lignin stream must be vaporized at a net BTU cost for the stream (stm 601B). Water vaporized is

43,969 kg/hr = 369,337 metric ton/year is vaporized at 1,100 BTU/lb loss = (107) MM BTU/hr

The remaining 19,809 kg/hr of stream 601B has 24,251 BTU/kg value = 480 MM BTU/hr

Total heating value from stream 601A is 374 MM BTU/hr

Gross Lignin Value/year = $7,848,926

Transport Cost = $7,848,926

Net Lignin Value = $0

METHANE

The digester produces 85% methane @ 353 kg/hr (stm 615) 44,332 BTU/kg CH4

Total heating value from Methane is 16 MM BTU/hr

methane is used in the DDG dryers and based on BTU value of $2.50 MM BTU

METHANE Value/year = $328,822

DIGESTER SLUDGE

The digester produces (stm 623) 0 kg/hr of sludge as fuel = 2,254 BTU/lb

based on 9,845 btu/lb biomass and 70% water in the sludge. = 4,969 BTU/kg

Total heating value from sludge is 0.00 MM BTU/hr

SLUDGE Value/year = $0

Sale of methane and lignin, based on BTU value is **$328,822 per year**

Total projected facility sales would be $28,649,558 per year

High Plains Corp.
York, NE Co-located
Stover-to-Ethanol Plant

CAPITAL INVESTMENT ASSUMPTIONS

1) Total capital investment

Civil Structural			1,500,000	
Area 100			6,146,434	
Area 200			14,955,166	
Area 300			4,028,307	
Area 307			3,714,334	
Area 400			8,676,000	
Area 500			7,515,486	
Area 600			9,824,251	
Area 700			273,557	
Area 800			3,684,612	
Area 900			2,236,491	
Fixed Capital			$62,554,640	
INDIRECTS	Prorateable	3.5%	$2,189,412	
	Process Development	2.0%	$1,251,093	
	Field Expense	8.0%	$5,004,371	
	Home Office Constr. Fee	12.0%	$7,506,557	
	Contingency	10.0%	$6,255,464	
	Start-up, Permits, Fees	3.0%	$1,876,639	
Working Capital per estimate			$1,590,867	1 mos Raw matls. + O&M
	Total Plant Cost		$88,229,044	
FEDERAL & STATE GRANTS		10%	($8,822,904)	
	Net Capital Investment		**$79,406,139**	

OPERATING COST ASSUMPTIONS 8,400 hr/yr

Utilities (Rates based on 25,746,124 gal/yr produced)

	Amount/hr	*Units*	*$/unit*	*Cost /hr.*	**Total Cost /yr**
*Electricity	12,893	Kw-hr	$0.035	$451	$3,790,636
Well water	79,972	kg	$0.000	$0	$0
*Wastewater	39,119	kg	$0.00026	$10	$86,808
*Gypsum waste disposal	1,137	kg	$0.0364	$41	$347,327
		mTon	$1.103	$0	$0
Total Utilities				$503	$4,224,771

Quoted by High Plains

High Plains Corp.
York, NE Co-located
Stover-to-Ethanol Plant

Raw Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Corn Stover DRY (stm 101 less water)	37,500	kg	$0.016	$597.41	$5,018,284
*Sulfuric Acid (stm 710)	860	kg	$0.100	$86.26	$724,592
*Calcium Hydroxide (Lime stm 227)	337	kg	$0.293	$98.70	$829,039
*Ammonia (stm 717)	445	kg	$0.162	$72.07	$605,374
Corn Steep Liquor (stm 735)	859	kg	$0.051	$43.80	$367,909
Nutrients (stm 415)	60	kg	$0.291	$17.48	$146,846
Purchased Cellulase	0	kg	$3.000	$0.00	$0
*Natural Gasoline (stm 701)	391	kg	$0.155	$60.36	$506,988
*Rolling Stock Gasoline	79	kg	$0.155	$12.32	$103,470
*WWT Chemicals	5	kg	$2.237	$11.98	$100,603
*CW Chemicals	17	kg	$1.428	$24.38	$204,791
*BFW Chemicals	73.8	kg	$0.226	$16.65	$139,833
*Boiler Fuel (stm 813)	190	Mbtu	$2.500	$476.07	$3,998,989
Total Raw Materials				$1,517	$12,746,718

**** Quoted by High Plains***

Processing Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
*Antifoam (Corn Oil)	79	kg	$0.304	$24	$200,961
Total Processing Materials				$24	$200,961

**** Quoted by High Plains***

Operations and Maintenance Costs - DRY HANDLING (area 100)	each/day		wage	hr/day each	Total Cost /yr.
*Supervisors	0.5	$	20.00	12	$43,800
*Operators	2.0	$	16.00	12	$140,160
*Laborers	8.0	$	16.00	12	$560,640
*Maintenance	2.0	$	16.00	12	$140,160
Operations and Maintenance Costs - HYDROLYSIS/FERMENTATION (area 200, 300, 400, 500, 600)					
*Supervisors	1.0	$	20.00	12	$87,600
*Operators	9.0	$	16.00	8	$420,480
*Laborers	4.0	$	16.00	8	$186,880
*Technicians (Includes Lab.)	3.0	$	16.00	8	$140,160
*Maintenance	3.0	$	16.00	8	$140,160
Operations and Maintenance Costs - Utilities (area 700, 800, 900)					
*Supervisors	0.5	$	20.00	12	$21,900
*Operators	3.0	$	16.00	8	$70,080
*Laborers	1.0	$	16.00	8	$23,360
*Technicians	1.0	$	16.00	8	$23,360
*Maintenance	2.0	$	16.00	8	$46,720

**** Quoted by High Plains*** Standard HPY shifts are 12 hours.

Total Operations and maintenance labor costs $2,045,460

High Plains Corp.
York, NE Co-located
Stover-to-Ethanol Plant

Other Operations and Maintenance Costs			
Payroll Overhead	35% of operating labor	$	715,911
Maintenance Costs	2% of plant cost	$	1,251,093
Operating Supplies	0.25% of plant cost	$	156,387
Environmental	0.50% of plant cost	$	312,773
Local Taxes	1% of plant cost	$	625,546
Insurance	0.50% of plant cost	$	312,773
Overhead Costs	40% of labor, supervision,maint cost	$	818,184
Administrative Costs	1% of annual sales (less tax credits)	$	105,559
Distribution and Sales	0.5% of annual sales (less tax credits)	$	-
Total O&M Costs			$6,343,686

OTHER MODEL ASSUMPTIONS

Average prevailing market price of fuel grade ETOH:		$0.37	per kg
Assumes renewal of the ethanol excise tax credit of $.54 per gallon	$	1.10	per gallon
and the small producer tax credit of $.10 per gallon through the year 2007			
*Value of CO_2 produced	$	4.13	per metric ton
*Price for Electricity		$0.035	per KWhr
*Gas price per million BTU		$2.500	per MM BTU

			68%	Dry matter
Corn Stover feedstock cost- **dry basis/short ton**	$	14.45	$0.016	per kg
			$15.93	per metric ton

Plant on-stream factor	0.959	
Plant operating hours per year	8400	
Depreciable Life of Capital Equipment	15	years
Average annual commodity escalation rate:	3.0%	
Average annual cost escalation rate:	3.0%	

Quoted by High Plains

1. There are no land acquisiton costs included.
2. There are no off site costs included (e.g. public road improvements, extensions of power, water, telephone services)
3. There is a source of qualified construction personnel within daily driving distance of the site
4. There exist adequate roads and rail roads to allow equipment delivery.
5. The costs for air and water permits are not included.
6. Soils are adequate for conventional foundation designs.

EL ENZYMATIC HYDROLYSIS - PRO FORMA

A 1/26/00

E 1: Produce Fuel Grade Ethanol

ital Investment:		month1	month2	month3	month4	month5	month6	month7	month8	month9	month10	month11	month12	month13	month14	month15	month16	month17	month18	TOTAL
otal Fixed Capital Cost		$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$4,000,000	$11,406,139	$79,406,139
onstruction Financing & Fees @10%			33,333	66,667	100,000	133,333	166,667	200,000	233,333	266,667	300,000	333,333	366,667	400,000	433,333	466,667	500,000	533,333	266,667	4,800,000
oan Origination Fee @	2.0%	1,588,123																		1,588,123
egal Fees		40,000																	0	40,000
uilder's All Risk/General Liability		50,000																	0	50,000
orking Capital (Financed)																			0	0
Total Capital Investment Required		$5,678,123	$4,033,333	$4,066,667	$4,100,000	$4,133,333	$4,166,667	$4,200,000	$4,233,333	$4,266,667	$4,300,000	$4,333,333	$4,366,667	$4,400,000	$4,433,333	$4,466,667	$4,500,000	$4,533,333	$11,672,806	$85,884,262
					11,000															

rating Projection:	Year 1: 1999 / 2000	Year 2: 2000/2001	Year 3: 2001/2002	Year 4: 2002/2003	Year 5: 2003/2004	Year 6: 2004/2005	Year 7: 2005/2006	Year 8: 2006/2007	Year 9: 2007/2008	Year 10: 2008/2009	Year 11: 2006 / 2007	Year 12: 2007 / 2008	Year 13: 2008 / 2009	Year 14: 2009 / 2009	Year 15: 2010 / 2011	Year 16: 2011 / 2012	Year 17: 2012 / 2013	Year 18: 2013 / 2014	Year 19: 2014 / 2015	Year 20: 2015 / 2016
l of fuel grade ethanol produced	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124	25,746,124
ontract sale price per gallon	$1.100	$1.133	$1.167	$1.202	$1.238	$1.275	$1.313	$1.353	$1.393	$1.435	$1.478	$1.523	$1.568	$1.615	$1.664	$1.714	$1.765	$1.818	$1.873	$1.929
Gross Annual Revenue	$28,320,736	$29,170,358	$30,045,469	$30,946,833	$31,875,238	$32,831,495	$33,816,440	$34,830,933	$35,875,861	$36,952,137	$38,060,701	$39,202,522	$40,378,598	$41,589,956	$42,837,655	$44,122,784	$45,446,468	$46,809,862	$48,214,158	$49,660,582
mall Ethanol Producer Tax Credit @ $0.0000 per gallon	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Total projected ethanol sales and credit	$28,320,736	$29,170,358	$30,045,469	$30,946,833	$31,875,238	$32,831,495	$33,816,440	$34,830,933	$35,875,861	$36,952,137	$38,060,701	$39,202,522	$40,378,598	$41,589,956	$42,837,655	$44,122,784	$45,446,468	$46,809,862	$48,214,158	$49,660,582
alue of electricity	$0.035	$0.036	$0.037	$0.038	$0.039	$0.041	$0.042	$0.043	$0.044	$0.046	$0.047	$0.048	$0.050	$0.051	$0.053	$0.055	$0.056	$0.058	$0.060	$0.061
Gross Annual Co-Product Revenue	$328,822	$338,687	$348,847	$359,313	$370,092	$381,195	$392,631	$404,410	$416,542	$429,038	$441,909	$455,167	$468,822	$482,886	$497,373	$512,294	$527,663	$543,493	$559,797	$576,591
Gross Sales and Credit	$28,649,558	$29,509,045	$30,394,316	$31,306,146	$32,245,330	$33,212,690	$34,209,071	$35,235,343	$36,292,403	$37,381,175	$38,502,611	$39,657,689	$40,847,420	$42,072,842	$43,335,027	$44,635,078	$45,974,131	$47,353,354	$48,773,955	$50,237,174
perating Expenses:																				
Utilities	4,224,771	4,351,514	4,482,059	4,616,521	4,755,017	4,897,667	5,044,597	5,195,935	5,351,813	5,512,368	5,677,739	5,848,071	6,023,513	6,204,218	6,390,345	6,582,055	6,779,517	6,982,902	7,192,389	7,408,161
Raw Materials	12,746,716	13,129,120	13,522,993	13,928,683	14,346,544	14,776,940	15,220,248	15,676,856	16,147,161	16,631,576	17,130,524	17,644,439	18,173,772	18,718,986	19,280,555	19,858,972	20,454,741	21,068,383	21,700,435	22,351,448
Processing Materials	200,961	206,990	213,200	219,596	226,184	232,969	239,958	247,157	254,572	262,209	270,075	278,177	286,523	295,119	303,972	313,091	322,484	332,158	342,123	352,387
Operation & Maintenance	6,343,686	6,533,997	6,730,017	6,931,917	7,139,875	7,354,071	7,574,693	7,801,934	8,035,992	8,277,072	8,525,384	8,781,145	9,044,580	9,315,917	9,595,395	9,883,257	10,179,754	10,485,147	10,799,701	11,123,692
Property Tax @ 0.50% Book Value	429,421	402,953	376,484	350,015	323,546	297,078	270,609	244,140	217,672	191,203	164,734	138,265	111,797	85,328	58,859	58,859	58,859	58,859	58,859	58,859
Depreciation	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	5,293,743	0	0	0	0	0
Total Operating Expense	$29,239,301	$29,918,316	$30,618,496	$31,340,475	$32,084,908	$32,852,468	$33,643,849	$34,459,765	$35,300,953	$36,168,170	$37,062,198	$37,983,841	$38,933,927	$39,913,310	$40,922,869	$36,696,234	$37,795,355	$38,927,450	$40,093,508	$41,294,547
Net Operating Income	($589,742)	($409,271)	($224,180)	($34,330)	$160,422	$360,222	$565,222	$775,578	$991,450	$1,213,005	$1,440,412	$1,673,848	$1,913,492	$2,159,532	$2,412,159	$7,938,844	$8,178,775	$8,425,904	$8,680,447	$8,942,626
Net Operating Cash Flow	$4,704,000	$4,884,471	$5,069,563	$5,259,413	$5,454,165	$5,653,965	$5,858,965	$6,069,321	$6,285,193	$6,506,748	$6,734,155	$6,967,590	$7,207,235	$7,453,274	$7,705,901	$7,938,844	$8,178,775	$8,425,904	$8,680,447	$8,942,626

E 1: Hypothetical Financing Scenarios:

E 1A: 100% Debt Financing

mortization 15 yrs terest Rate 7.00%	Year 1: 1999 / 2000	Year 2: 2000/2001	Year 3: 2001/2002	Year 4: 2002/2003	Year 5: 2003/2004	Year 6: 2004/2005	Year 7: 2005/2006	Year 8: 2006/2007	Year 9: 2007/2008	Year 10: 2008/2009	Year 11: 2006 / 2007	Year 12: 2007 / 2008	Year 13: 2008 / 2009	Year 14: 2009 / 2009	Year 15: 2010 / 2011	Year 16: 2011 / 2012	Year 17: 2012 / 2013	Year 18: 2013 / 2014	Year 19: 2014 / 2015	Year 20: 2015 / 2016
et Operating Cash Flow *(from above)*	4,704,000	4,884,471	5,069,563	5,259,413	5,454,165	5,653,965	5,858,965	6,069,321	6,285,193	6,506,748	6,734,155	6,967,590	7,207,235	7,453,274	7,705,901	7,938,844	8,178,775	8,425,904	8,680,447	8,942,626
Debt Interest	6,011,898	5,772,657	5,516,669	5,242,762	4,949,681	4,636,084	4,300,536	3,941,500	3,557,330	3,146,269	2,706,434	2,235,810	1,732,243	1,193,426	616,892	0	0	0	0	0
Debt Principal	3,417,732	3,656,973	3,912,961	4,186,869	4,479,949	4,793,546	5,129,094	5,488,131	5,872,300	6,283,361	6,723,196	7,193,820	7,697,387	8,236,204	8,812,739	(0)	(0)	(0)	(0)	(0)
Total Debt Service	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	9,429,630	0	0	0	0	0
Net Cash Flow after Debt Service	(4,725,630)	(4,545,159)	(4,360,067)	(4,170,217)	(3,975,466)	(3,775,666)	(3,570,666)	(3,360,310)	(3,144,437)	(2,922,883)	(2,695,475)	(2,462,040)	(2,222,396)	(1,976,356)	(1,723,729)	7,938,844	8,178,775	8,425,904	8,680,447	8,942,626
Debt Service Coverage Ratio	0.50	0.52	0.54	0.56	0.58	0.60	0.62	0.64	0.67	0.69										
Total Pre-tax Net Cash Flow *(20 yrs)*	($7,463,900)																			

E 1B: 100% Cash Financing

	Year 0:	Year 1:	Year 2:	Year 3:	Year 4:	Year 5:	Year 6:	Year 7:	Year 8:	Year 9:	Year 10:	Year 11:	Year 12:	Year 13:	Year 14:	Year 15:	Year 16:	Year 17:	Year 18:	Year 19:	Year 20:
et Cash Flow	(85,884,262)	4,704,000	4,884,471	5,069,563	5,259,413	5,454,165	5,653,965	5,858,965	6,069,321	6,285,193	6,506,748	6,734,155	6,967,590	7,207,235	7,453,274	7,705,901	7,938,844	8,178,775	8,425,904	8,680,447	8,942,626
tal Pre-tax Net Cash Flow *(20 yrs)*		$48,096,293																			
IRR @ 100% CASH		4.15%																			
yback Period *(Pre-tax; undiscounted)*		(18.3) years																			

SE 1C: Combined Equity & Debt Financing

quity Portion	**25.00%**	$21,471,066	Amortization	**15**	yrs
ebt Portion	**75.00%**	$64,413,197	Interest Rate	**7.00%**	

	Year 0: 1997/1998	*Year 1: 1999 / 2000*	*Year 2: 2000/2001*	*Year 3: 2001/2002*	*Year 4: 2002/2003*	*Year 5: 2003/2004*	*Year 6: 2004/2005*	*Year 7: 2005/2006*	*Year 8: 2006/2007*	*Year 9: 2007/2008*	*Year 10: 2008/2009*
et Operating Cash Flow	0	4,704,000	4,884,471	5,069,563	5,259,413	5,454,165	5,653,965	5,858,965	6,069,321	6,285,193	6,506,748
Debt Interest		4,508,924	4,329,493	4,137,502	3,932,071	3,712,261	3,477,063	3,225,402	2,956,125	2,667,998	2,359,702
Debt Principal		2,563,299	2,742,730	2,934,721	3,140,151	3,359,962	3,595,159	3,846,821	4,116,098	4,404,225	4,712,521
Total Debt Service		7,072,223	7,072,223	7,072,223	7,072,223	7,072,223	7,072,223	7,072,223	7,072,223	7,072,223	7,072,223
Net Cash Flow	(21,471,066)	(2,368,223)	(2,187,751)	(2,002,660)	(1,812,810)	(1,618,058)	(1,418,258)	(1,213,258)	(1,002,902)	(787,030)	(565,475)
ebt Service Coverage Ratio		*0.67*	*0.69*	*0.72*	*0.74*	*0.77*	*0.80*	*0.83*	*0.86*	*0.89*	*0.92*

	Year 11: 2006 / 2007	*Year 12: 2007 / 2008*	*Year 13: 2008 / 2009*	*Year 14: 2009 / 2009*	*Year 15: 2010 / 2011*	*Year 16: 2011 / 2012*	*Year 17: 2012 / 2013*	*Year 18: 2013 / 2014*	*Year 19: 2014 / 2015*	*Year 20: 2015 / 2016*
et Operating Cash Flow	6,734,155	6,967,590	7,207,235	7,453,274	7,705,901	7,938,844	8,178,775	8,425,904	8,680,447	8,942,626
Debt Interest	2,029,826	1,676,858	1,299,182	895,070	462,669	0	0	0	0	0
Debt Principal	5,042,397	5,395,365	5,773,040	6,177,153	6,609,554	(0)	(0)	(0)	(0)	(0)
Total Debt Service	7,072,223	7,072,223	7,072,223	7,072,223	7,072,223	0	0	0	0	0
Net Cash Flow	(338,068)	(104,632)	135,012	381,052	633,678	7,938,844	8,178,775	8,425,904	8,680,447	8,942,626
ebt Service Coverage Ratio	*0.95*	*0.99*	*1.02*	*1.05*	*1.09*	*#DIV/0!*	*#DIV/0!*	*#DIV/0!*	*#DIV/0!*	*#DIV/0!*

otal Pre-tax Net Cash Flow *(20 yrs)* $6,426,149

ternal Rate of Return *(IRR Pre-Tax)* 1.0%

odified Internal Rate of Return *(MIRR Pre-tax)* 1.4% *(excludes any assumption of project terminal value)*

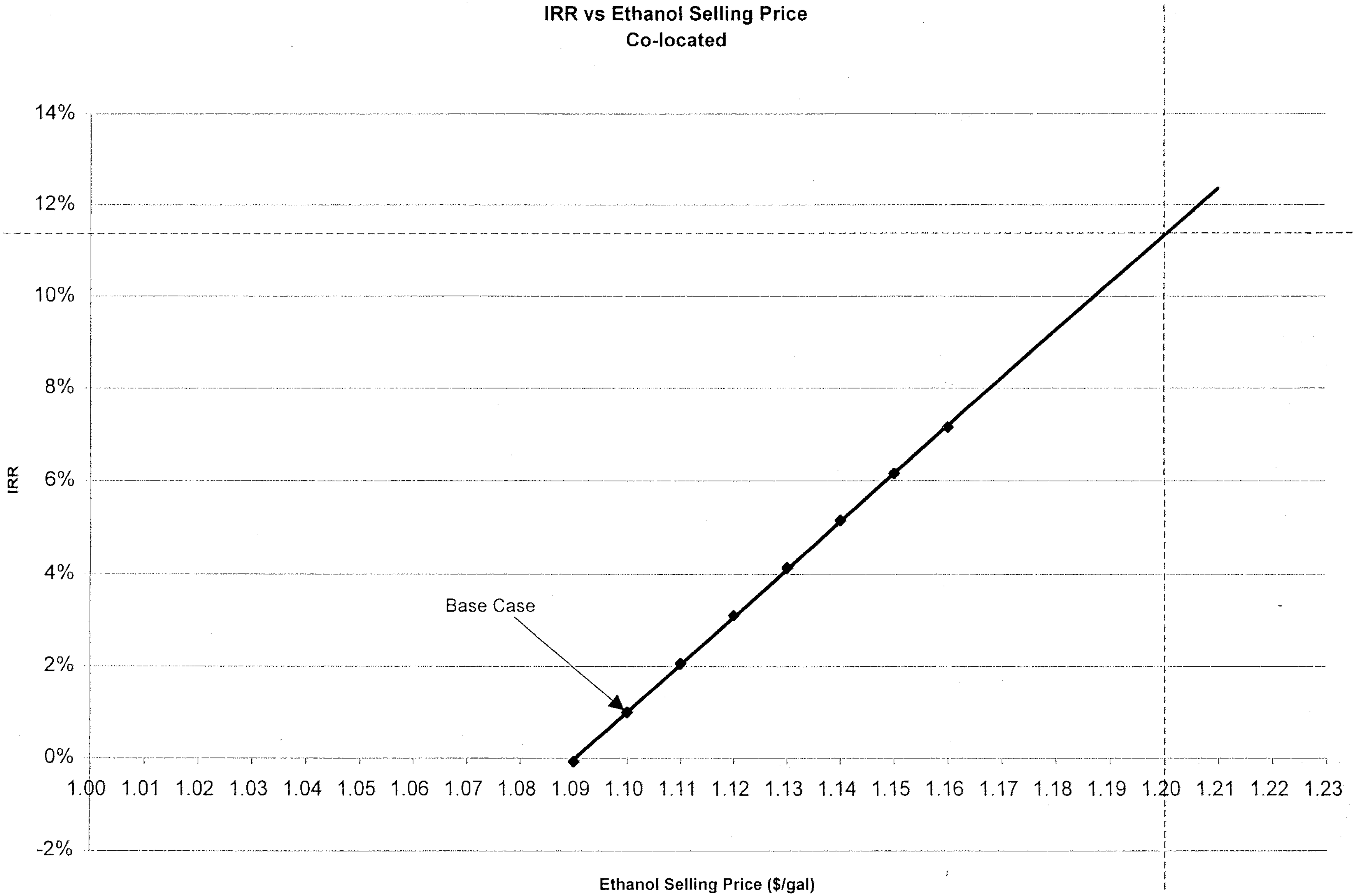
Sensitivity Analysis

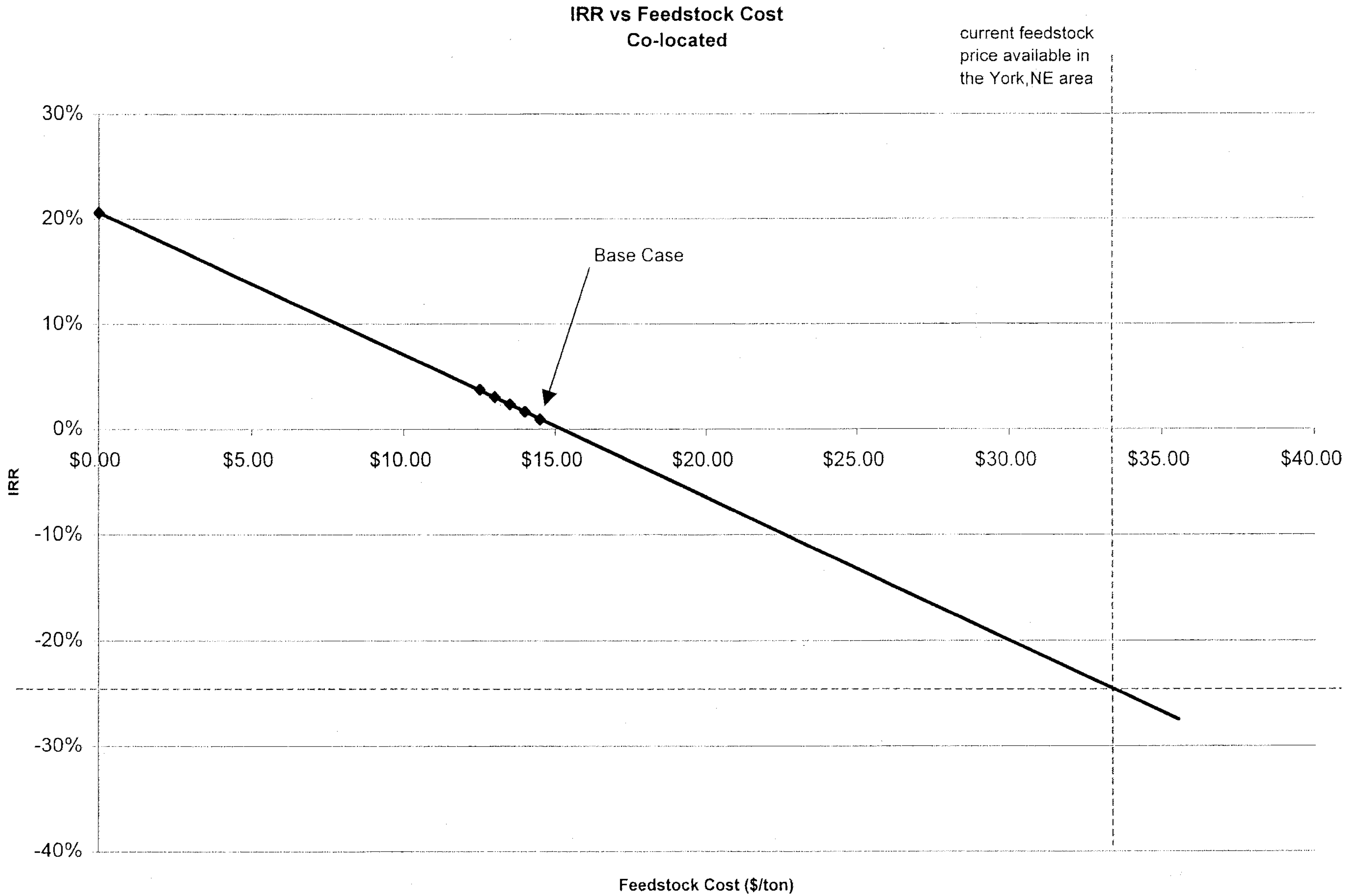
High Plains Corp.
York, NE Co-located
Stover-to-Ethanol Plant

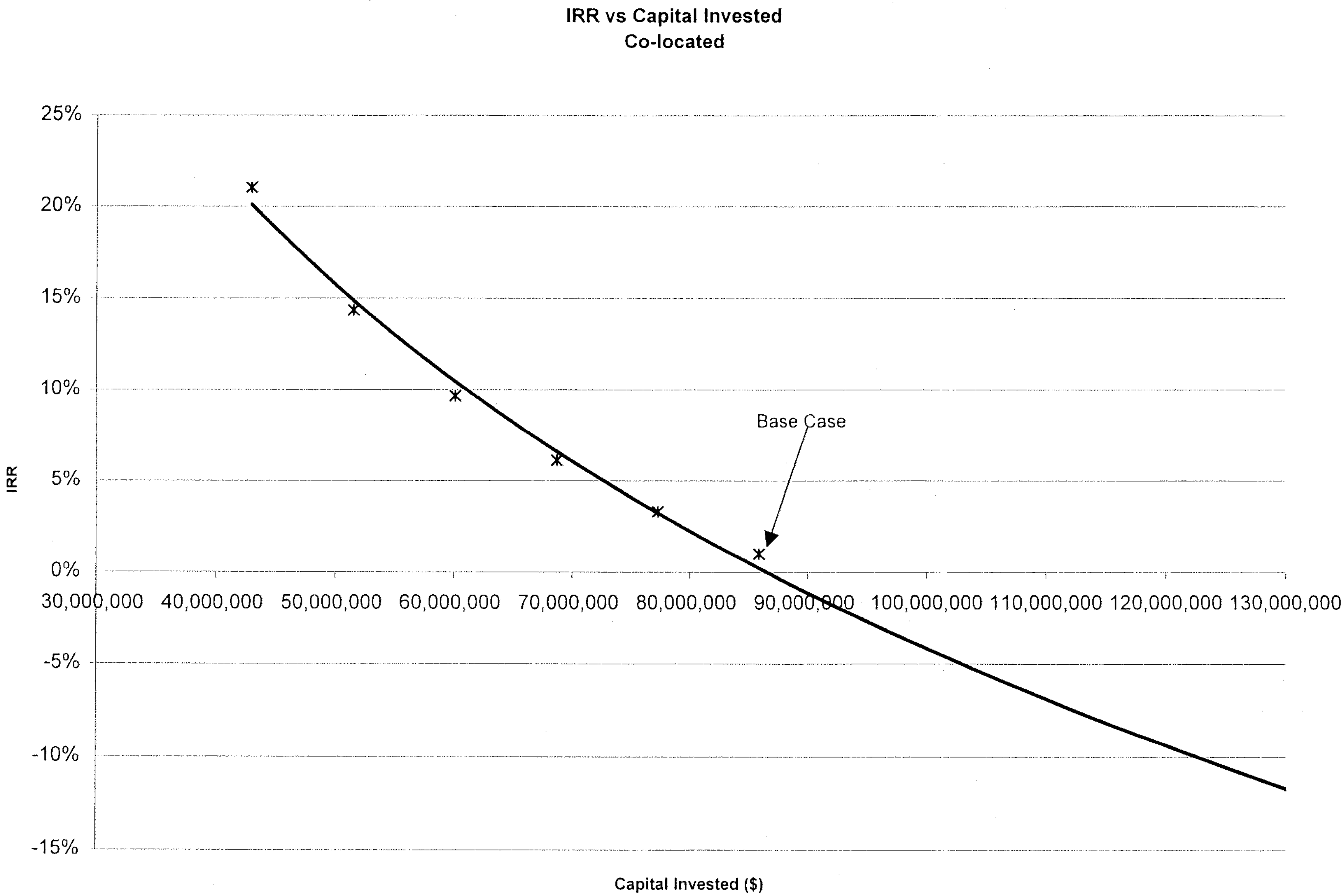
	20 YR. NET CASH FLOW $	RATE OF RETURN	FEED PRICE DRY $/TON	ETHANOL SALE $/GAL	
E1	(3,950,971)	-1%	14.450	1.085	
E2	(491,931)	0%	14.450	1.09	
E3	6,426,149	1%	14.450	1.10	BASE CASE
E4	13,344,229	2%	14.450	1.11	
E5	20,262,308	3%	14.450	1.12	
E6	27,180,388	4%	14.450	1.13	
E7	34,098,468	5%	14.450	1.14	
E8	41,016,548	6%	14.450	1.15	
E9	47,934,628	7%	14.450	1.16	
F1	141,269,327	21%	0.000	1.10	
F2	24,622,979	4%	12.500	1.10	
F3	19,957,125	3%	13.000	1.10	
F4	15,291,271	2%	13.500	1.10	
F5	10,625,417	2%	14.000	1.10	
F6	5,959,563	1%	14.500	1.10	

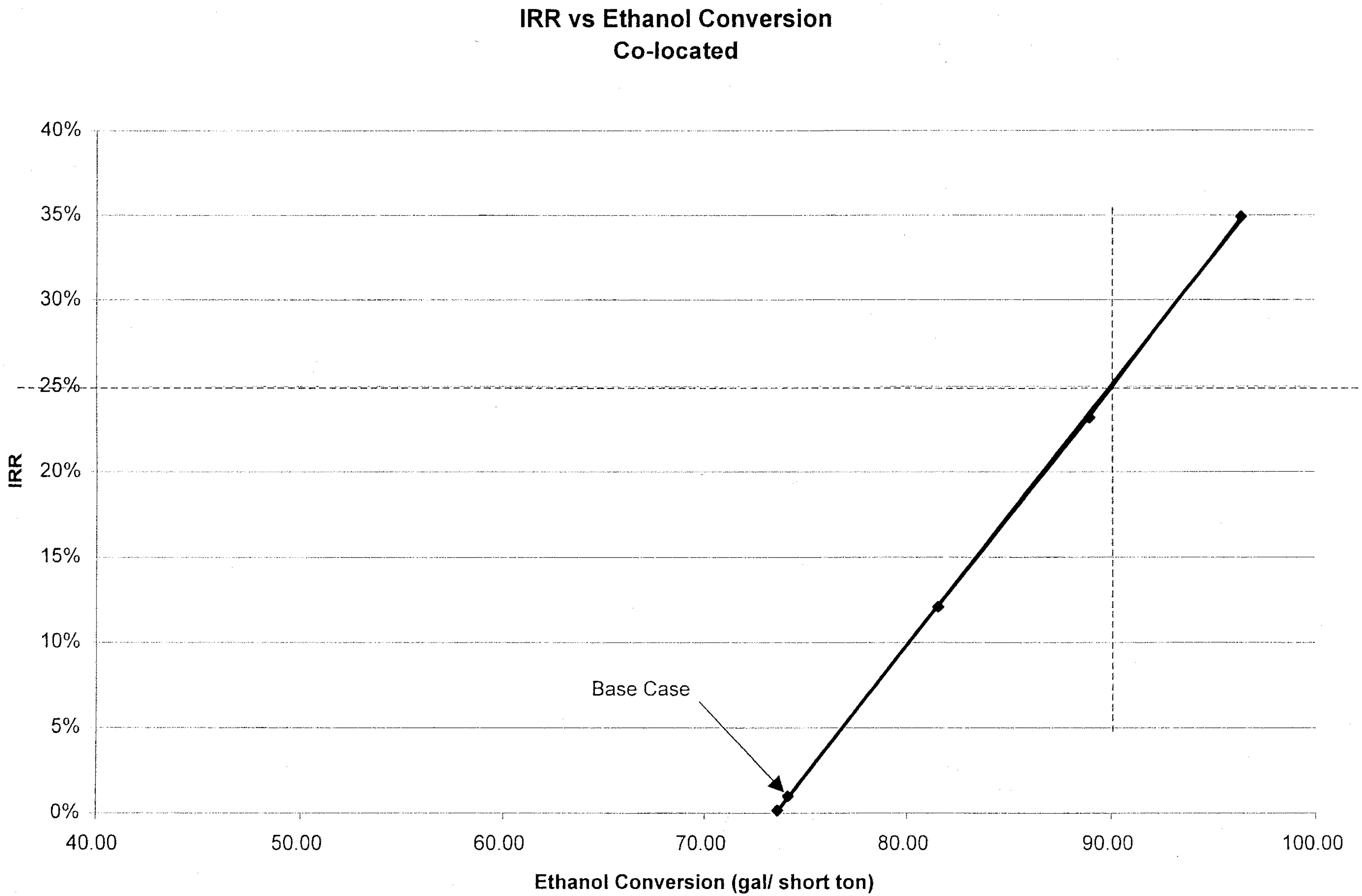
					%	CAPITAL INVEST	$/gal of capacity	
cap1	72,181,553	21%	14.450	1.10	50%	42,942,131	$ 1.67	
cap2	59,030,472	14%	14.450	1.10	60%	51,530,557	$ 2.00	
cap3	45,879,391	10%	14.450	1.10	70%	60,118,983	$ 2.34	
cap4	32,728,310	6%	14.450	1.10	80%	68,707,410	$ 2.67	
cap5	19,577,229	3%	14.450	1.10	90%	77,295,836	$ 3.00	
cap6	6,426,149	1%	14.450	1.10	100%	85,884,262	$ 3.34	BASE CASE
cap7	(6,724,932)	#NUM!	14.450	1.10	110%	94,472,688	$ 3.67	
cap8	(19,876,013)	#NUM!	14.450	1.10	120%	103,061,115	$ 4.00	
cap9	(33,027,094)	#DIV/0!	14.450	1.10	130%	111,649,541	$ 4.34	
cap10	(46,178,175)	#DIV/0!	14.450	1.10	140%	120,237,967	$ 4.67	
cap11	(59,329,255)	#DIV/0!	14.450	1.10	150%	128,826,393	$ 5.00	

					gal per short ton	Ethanol Produced		
p1	(374,068,245)	#DIV/0!	14.450	1.10	37.07	12,873,062	50%	
p2	(297,969,366)	#DIV/0!	14.450	1.10	44.49	15,447,674	60%	
p3	(221,870,488)	#DIV/0!	14.450	1.10	51.90	18,022,287	70%	
p4	(145,771,609)	#DIV/0!	14.450	1.10	59.32	20,596,899	80%	
p5	1,099,227	0%	14.450	1.10	73.63	25,565,901	99.30%	
p6	6,426,149	1%	14.450	1.10	74.15	25,746,124	100%	BASE CASE
p7	82,525,027	12%	14.450	1.10	81.56	28,320,736	110%	
p8	158,623,906	23%	14.450	1.10	88.98	30,895,349	120%	
p9	234,722,785	35%	14.450	1.10	96.39	33,469,961	130%	


IRR vs Ethanol Selling Price
Co-located
14%
12%
10%
8%
6%
4%
2%
0%
-2%
IRR
Base Case
1.00 1.01 1.02 1.03 1.04 1.05 1.06 1.07 1.08 1.09 1.10 1.11 1.12 1.13 1.14 1.15 1.16 1.17 1.18 1.19 1.20 1.21 1.22 1.23
Ethanol Selling Price ($/gal)


IRR vs Feedstock Cost
Co-located
current feedstock price available in the York,NE area
Base Case
30%
20%
10%
0%
-10%
-20%
-30%
-40%
IRR
$0.00
$5.00
$10.00
$15.00
$20.00
$25.00
$30.00
$35.00
$40.00
Feedstock Cost ($/ton)


IRR vs Capital Invested
Co-located
25%
20%
15%
10%
5%
0%
-5%
-10%
-15%
IRR
30,000,000
40,000,000
50,000,000
60,000,000
70,000,000
80,000,000
90,000,000
100,000,000
110,000,000
120,000,000
130,000,000
Base Case
Capital Invested ($)


IRR vs Ethanol Conversion
Co-located
40%
35%
30%
25%
20%
15%
10%
5%
0%
IRR
Base Case
40.00
50.00
60.00
70.00
80.00
90.00
100.00
Ethanol Conversion (gal/ short ton)

Comparison of Cellulase Sources

Comparison of On-site cellulase production methods

Comparison of On-Site Cellulase Production via Pure Vision Technology and NREL Reference Model

		NREL* M FPU required/yr**	difference	Pure Vision M FPU required/yr
		1,446,984	(50,708)	1,497,692
ng Projection:				
of fuel grade ethanol produced		$ 25,434,849	$ (311,275)	$ 25,746,124
itract sale price per gallon		$ 1	$ -	$ 1
Gross Annual Revenue		$ 27,978,334	$ (342,402)	$ 28,320,736
all Ethanol Producer Tax Credit				
@ $ - per gallon		$ -		$ -
Total projected ethanol sales and credit		$ 27,978,334	$ (342,402)	$ 28,320,736
Gross Annual Co-Product Revenue		$ 328,822	$ -	$ 328,822
Gross Sales and Credit		$ 28,307,156	$ (342,402)	$ 28,649,558
rating Expenses:				
tilities		$ 4,792,171	$ 567,400	$ 4,224,771
aw Materials		$ 12,843,241	$ 96,523	$ 12,746,718
rocessing Materials		$ 267,948	$ 66,987	$ 200,961
peration & Maintenance		$ 6,414,114	$ 70,428	$ 6,343,686
roperty Tax @ 0.50% Book Value		$ 486,736	$ 57,315	$ 429,421
epreciation		$ 6,038,644	$ 744,902	$ 5,293,743
Total Operating Expense		$ 30,842,855	$ 1,603,554	$ 29,239,301
et Operating Income		$ (2,535,699)	$ (1,945,956)	$ (589,742)
›t Operating Cash Flow		$ 3,502,945	$ (1,201,055)	$ 4,704,000
enzyme cost (cost of production calculated in "$per lb. calcs.") divided by lbs. per year flow rate from mass balance.	$/lb	$ 0.027		$ 0.020
enzyme cost (cost of production calculated in "$per lb. calcs.") divided by million FPU per year required.	**$/MFPU**	**$ 4.60**		**$ 3.32**

Annual Savings Using PureVision On-Site Enzyme Production	
OVER REFERENCE MODEL: $	**1,201,055**

* 45% scale factor applied, SHCF

* * MFPU = million FPU

JDY MODEL WITH REFRENCE MODEL CELLULASE PRODUCTION FOR COMPARISON OF ON-SITE ENZYME PRODUCTION VS. PURCHASE

LOSS OF ETOH PRODUCTION POSSIBLE: **111 kg/hr**

A

10/27/99

ENZYMATIC HYDROLYSIS - PRO FORMA

/ing Assumptions & Input Variables

RRENT SITUATION:

The Pro Forma models an Enzymatic Hydrolysis Ethanol plant using corn stover as the feed stock.

ETHANOL

The plant will convert corn stover to fuel grade ethanol utilizing enzymatic hydrolosis.

Corn stover feed rate of 71,977 kg/hr (str 101), produce estimated total output in

equivalent kilograms of fuel grade ETOH 9,041 kg/hr. = 75,942,299 kg / year (str 515)

gal./short ton= 73.3 3,028 gal/hr = 25,434,849 gal / year

gal./metric ton 80.7

Increase to current York yearly production: 69%

The model assumes renewal of the ethanol excise tax credit of $.54 per gallon to the blender

and **NOT** the small producer tax credit of $.10 per gallon through the year 2015 for a total ethanol value of

$1.10 per gallon or **$0.37** per kg and **$ 27,978,334 per year TOTAL Ethanol sales**

CARBON DIOXIDE

Currently, carbon dioxide from the High Plains York fermentations is sold to a CO_2 compression company.

Diverting the CO_2 (stm 550) from the stover plant into this stream for sale as opposed to the atmosphere provides

110,749 kg/hr = **930,294** ton / year with a value of $ 4.13 per metric ton

WITH THIS PROFORMA NO CO_2 IS SOLD. **CO_2 Value/year = $0**

LIGNIN

A Lignin co-product is produced and sold as combustion fuel material. A total amount of lignin in the stream (stm 601B) is

63,778 kg/hr = **535,734** metric ton / year is produced from the process.

The water in the lignin stream must be vaporized at a net BTU cost for the stream (stm 601B). Water vaporized is

43,969 kg/hr = **369,337** metric ton/year is vaporized at 1,100 BTU/lb loss = (107) MM BTU/hr

The remaining **19,809** kg/hr of stream 601B has **24,251** BTU/kg value = **480** MM BTU/hr

Total heating value from stream 601A is 374 MM BTU/hr

Gross Lignin Value/year = $7,848,926

Transport Cost = $7,848,926

Net Lignin Value = $0

METHANE

The digester produces 85% methane @ 353 kg/hr (stm 615) **44,332** BTU/kg CH4

Total heating value from Methane is 16 MM BTU/hr

methane is used in the DDG dryers and based on BTU value of $2.50 MM BTU

METHANE Value/year = $328,822

DIGESTER SLUDGE

The digester produces (stm 623) 0 kg/hr of sludge as fuel = **2,254** BTU/lb

based on 9,845 btu/lb biomass and 70% water in the sludge. = **4,969** BTU/kg

Total heating value from sludge is 0.00 MM BTU/hr

SLUDGE Value/year = $0

Sale of methane and lignin, based on BTU value is **$328,822 per year**

Total projected facility sales would be $28,307,156 per year

APITAL INVESTMENT ASSUMPTIONS

Total capital investment			
Civil Structural		1,500,000	
Area 100		6,146,434	
Area 200		14,955,166	
Area 300		4,028,307	
Area 307		3,714,334	
Area 400		10,353,995	
Area 500		7,515,486	
Area 600		9,824,251	
Area 700		282,716	
Area 800		3,684,612	
Area 900		2,236,491	
Fixed Capital		$64,241,793	
INDIRECTS Prorateable	3.5%	$2,248,463	
Process Development	2.0%	$1,284,836	
Field Expense	8.0%	$5,139,343	
Home Office Constr. Fee	12.0%	$7,709,015	
Contingency	10.0%	$6,424,179	
Start-up, Permits, Fees	3.0%	$1,927,254	
Working Capital per estimate		$1,604,780	1 mos Raw matls. + O&M
	Total Plant Cost	$90,579,663	
FEDERAL & STATE GRANTS	10%	($9,057,966)	
	Net Capital Investment	**$81,521,697**	

PERATING COST ASSUMPTIONS 8,400 hr/yr

Utilities (Rates based on 25,434,849 gal/yr produced)

	Amount/hr	*Units*	*$/unit*	*Cost /hr.*	**Total Cost /yr**
*Electricity	14,823	Kw-hr	$0.035	$519	$4,358,036
Well water	79,972	kg	$0.000	$0	$0
*Wastewater	39,119	kg	$0.00026	$10	$86,808
*Gypsum waste disposal	1,137	kg	$0.0364	$41	$347,327
		mTon	$1.103	$0	$0
Total Utilities				$570	$4,792,171

**** Quoted by High Plains***

Raw Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Corn Stover DRY (stm 101 less water)	37,500	kg	$0.016	$597.41	$5,018,284
*Sulfuric Acid (stm 710)	860	kg	$0.100	$86.26	$724,592
*Calcium Hydroxide (Lime stm 227)	337	kg	$0.293	$98.70	$829,039
*Ammonia (stm 717)	464	kg	$0.162	$75.17	$631,405
Corn Steep Liquor (stm 735)	909	kg	$0.051	$46.36	$389,452
Nutrients (stm 415)	80	kg	$0.291	$23.31	$195,794
Purchased Cellulase	0	kg	$3.000	$0.00	$0
*Natural Gasoline (stm 701)	391	kg	$0.155	$60.36	$506,988
*Rolling Stock Gasoline	79	kg	$0.155	$12.32	$103,470
*WWT Chemicals	5	kg	$2.237	$11.98	$100,603
*CW Chemicals	17	kg	$1.428	$24.38	$204,791
*BFW Chemicals	73.8	kg	$0.226	$16.65	$139,833
*Boiler Fuel (stm 813)	190	Mbtu	$2.500	$476.07	$3,998,989
Total Raw Materials				$1,529	$12,843,241

**** Quoted by High Plains***

Processing Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
*Antifoam (Corn Oil)	105	kg	$0.304	$32	$267,948
Total Processing Materials				$32	$267,948

**** Quoted by High Plains***

Operations and Maintenance Costs - DRY HANDLING (area 100	each/day		wage	hr/day each	Total Cost /yr.
*Supervisors	0.5	$	20.00	12	$43,800
*Operators	2.0	$	16.00	12	$140,160
*Laborers	8.0	$	16.00	12	$560,640
*Maintenance	2.0	$	16.00	12	$140,160
Operations and Maintenance Costs - HYDROLYSIS/FERMENTATION (area 200, 300, 400, 500, 600)					
*Supervisors	1.0	$	20.00	12	$87,600
*Operators	9.0	$	16.00	8	$420,480
*Laborers	4.0	$	16.00	8	$186,880
*Technicians (Includes Lab.)	3.0	$	16.00	8	$140,160
*Maintenance	3.0	$	16.00	8	$140,160
Operations and Maintenance Costs - Utilities (area 700, 800, 900)					
*Supervisors	0.5	$	20.00	12	$21,900
*Operators	3.0	$	16.00	8	$70,080
*Laborers	1.0	$	16.00	8	$23,360
*Technicians	1.0	$	16.00	8	$23,360
*Maintenance	2.0	$	16.00	8	$46,720

**** Quoted by High Plains*** Standard HPY shifts are 12 hours.

Total Operations and maintenance labor costs $2,045,460

Other Operations and Maintenance Costs

Payroll Overhead	35% of operating labor	$	715,911
Maintenance Costs	2% of plant cost	$	1,284,836
Operating Supplies	0.25% of plant cost	$	160,604
Environmental	0.50% of plant cost	$	321,209
Local Taxes	1% of plant cost	$	642,418
Insurance	0.50% of plant cost	$	321,209
Overhead Costs	40% of labor, supervision,maint cost	$	818,184
Administrative Costs	1% of annual sales (less tax credits)	$	104,283
Distribution and Sales	0.5% of annual sales (less tax credits)	$	-
Total O&M Costs			$6,414,114

THER MODEL ASSUMPTIONS

verage prevailing market price of fuel grade ETOH:	$0.37	per kg
ssumes renewal of the ethanol excise tax credit of $.54 per gallon	$ 1.10	per gallon
d the small producer tax credit of $.10 per gallon through the year 2007		
Value of CO_2 produced	$ 4.13	per metric ton
Price for Electricity	$ 0.035	per KWhr
Gas price per million BTU	$ 2.500	per MM BTU

		68%	Dry matter
orn Stover feedstock cost- **dry basis/short ton**	$ 14.45	$0.016	per kg
		$15.93	per metric ton

ant on-stream factor	0.959	
ant operating hours per year	8,400	
epreciable Life of Capital Equipment	15	years
verage annual commodity escalation rate:	3.0%	
verage annual cost escalation rate:	3.0%	

**** Quoted by High Plains***

There are no land acquisiton costs included.
There are no off site costs included (e.g. public road improvements, extensions of power, water, telephone services)
There is a source of qualified construction personnel within daily driving distance of the site
There exist adequate roads and rail roads to allow equipment delivery.
The costs for air and water permits are not included.
Soils are adequate for conventional foundation designs.

stimated Equipment Costs for Reference Model Scaled Down 45% with On-Site Enzyme Production and SHCF

all cells are automatically updated from file Equipa with the exception of red letter areas

Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Install Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description	3442 WORK	
C-101	1	0	Bale conveyor	AREA0100	154	170	1.11	$15,000	1999	$15,000	0.6	$15,927	1.5	$24,551	$ 15,927	wire mesh conveyor 60" wide 20' long	WC101	11.93
C-102	1	0	Radial Stacker Conveyor	AREA0100	154	170	1.11	$159,830	1999	$159,830	0.6	$169,708	1.5	$261,604	$ 169,708	16 degree, 36" x 200' radial stacker, 750 ton/hr, 75 HP	WC102	44.74
C-103	1	0	Breaker Infeed Belt	AREA0100	154	170	1.11	$49,500	1999	$49,500	0.6	$52,559	1.5	$81,020	$ 52,559	84" x 35' rubber belt cleated infeed conveyor, 10 HP, TEFC drive motor with guard	WC103	5.97
C-104	1	0	1st Shredder Conveyor	AREA0100	154	170	1.11	$25,650	1999	$25,650	0.6	$27,235	1.5	$41,983	$ 27,235	60" wide x 25' long, 10 HP, TEFC drive with guard	WC104	5.97
C-105	1	0	1st Infeed Belt	AREA0100	154	170	1.11	$38,500	1999	$38,500	0.6	$40,879	1.5	$63,015	$ 40,879	60" wide x 30' long, 10 HP, TEFC drive with guard	WC105	11.93
C-106	1	0	2nd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.5	$48,285	$ 31,323	48" wide x 20' long, 7.5 HP, TEFC drive with guard	WC106	4.47
C-107	1	0	2nd Infeed Belt	AREA0100	154	170	1.11	$27,500	1999	$27,500	0.6	$29,200	1.5	$45,011	$ 29,200	48" wide x 30' long, 5 HP, TEFC drive with guard	WC107	2.98
C-108	1	0	3rd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.5	$48,285	$ 31,323	48" wide x 20' long, 10 HP, TEFC drive with guard	WC108	5.97
C-109	1	0	Feed Screw Conveyor	AREA0100	225,140	562,850	2.50	$31,700	1997	$31,700	0.6	$54,932	1.5	$86,351	$ 56,018	14" dia. 250' long	WC109	53.75
M-101	2	0	Truck Scale	AREA0100	96	72	0.75	$10,000	1999	$20,000	0.6	$16,829	1.5	$25,244	$ 16,829	96 deliveries /scale/12hr		
M-102	1	0	Receiving Pad	AREA0100	250,000	250,000	1.00	$2,083,500	1999	$2,083,500	0.6	$2,083,500	1.0	$2,083,500	$ 2,083,500	250,000 ft2 concrete pad, 9" thick with drainage		
M-103	6	1	Front End Loader	AREA0100	159,948	159,948	1.00	$156,000	1998	$1,092,000	0.6	$1,092,000	1.2	$ 1,326,016	$ 1,105,013	run on gasoline		
M-104	3	0	Bale Breaker	AREA0100	154	170	1.11	$250,000	1999	$750,000	0.6	$796,352	1.2	$955,622	$ 796,352	30 HP each	WM104	53.69
M-105	1	0	Primary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.2	$135,444	$ 112,870	250 HP, 1200 rpm, hammermill	WM105	149.14
M-106	1	0	Secondary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.5	$169,304	$ 112,870	250 HP, 1200 rpm, hammermill	WM106	149.14
M-107	1	0	Shred Bunker	AREA0100	600,000	600,000	1.00	$700,000	1999	$700,000	0.6	$700,000	1.0	$700,000	$ 700,000	200x100x30ft bunker with three walls, 3 days shred storage		
M-108	1	0	Storm Runoff Pond	AREA0100	1,747,767	1,747,767	1.00	$51,198	1998	$51,198	0.6	$51,198	1.0	$51,198	$ 51,808	200 x 150 x 8 ft, 240,000ft3		
										weighted averages:	0.60		1.13					499.68
A100									Subtotal	$5,315,978		$5,418,705		$6,146,434	$5,433,414			
								2000lpd x .45 (current year cost with area weighted-average scale exponent applied)					1.3	$3,181,636	($2,964,798)	is installed cost savings		
											Cost Base Year =	1999						
A-201	1	0	In-line Sulfuric Acid Mixer	STRM0214	55,308	23,725	0.43	$1,900	1997	$1,900	0.48	$1,266	1.2	$1,585	$1,291	Static Mixer, 110 gpm total flow		
A-202	1	0	In-line NH3 Mixer	STRM0244	53,630	18,317	0.34	$1,500	1997	$1,500	0.48	$896	1.2	$1,122	$913	Static Mixer, 82 gpm total flow		
A-209	1	0	Overliming Tank Agitator	STRM0228	167,050	102,608	0.61	$19,800	1997	$19,800	0.51	$15,442	1.2	$19,345	$15,748	Top Mounted, 1800 rpm, 15 hp	WT209	8.39
A-224	1	0	Reacidification Tank Agitator	STRM0239	167,280	102,752	0.61	$65,200	1997	$65,200	0.51	$50,851	1.2	$63,702	$51,857	Top-Mounted, 1800 rpm, 54 hp	WT224	25.17
A-232	1	0	Reslurrying Tank Agitator	STRM0250	358,810	167,795	0.47	$36,000	1997	$36,000	0.51	$24,432	1.2	$30,606	$24,915	Top-Mounted, 1800 rpm, 25 hp	WT232	13.98
A-235	1	0	In-line Acidification Mixer	STRM0236	164,570	101,104	0.61	$2,600	1997	$2,600	0.48	$2,058	1.2	$2,578	$2,099	Static-Mixer, 440 gpm total flow		
C-201	1	0	Hydrolyzate Screw Conveyor	STRM0220	225,140	101,493	0.45	$59,400	1997	$59,400	0.78	$31,908	1.5	$50,158	$32,539	18" dia. 33' long, 3420 cfh max flow, 23 hp	WC201	13.72
C-202	1	0	Wash Solids Screw Conveyor	STRM0225	196,720	165,453	0.84	$23,700	1997	$23,700	1	$19,933	1.5	$31,334	$20,327	18" dia. 16' long, 3420 cfh max flow	WC202	16.70
C-225	1	0	Lime Solids Feeder			0		$3,900	1997	$3,900	1	$3,900	1.5	$6,131	$3,977	6" dia., 63 cfh, 3150 lb/hr max flow	WC225	0.15
H-200	1	0	Hydrolyzate Cooler	AREA0200	1,988	895	0.45	$45,000	1997	$45,000	0.51	$29,947	2.2	$66,543	$30,539	Fixed Tube Sheet, 900 sf, 20" dia. X 20' long		
H-201	1	1	Beer Column Feed Economizer	AREA0201	5,641	5,641	1.00	$139,350	1999	$278,700	0.68	$278,700	2.2	$607,278	$278,700	TEMA type AES shell and tube 5641 sf, 42" dia x 20' long		
M-202	1	0	Prehydrolysis Reactor	STRM0217	270,034	121,514	0.45	$12,461,841	1998	$12,461,841	0.78	$6,684,746	1.5	$10,146,612	$6,764,408	Vertical Screw, 10 min residence time	WM105	353.16
P-201	1	1	Sulfuric Acid Pump	STRM0710	1,647	414	0.25	$4,800	1997	$9,600	0.79	$3,228	2.8	$9,190	$3,291	2 gpm, 245 ft. head	WP201	0.40
P-209	1	1	Overlimed Hydrolyzate Pump	STRM0228	167,050	102,608	0.61	$10,700	1997	$21,400	0.79	$14,561	2.8	$41,458	$14,849	448 gpm, 150 ft. head	WP209	18.01
P-222	1	1	Filtered Hydrolyzate Pump	STRM0230	162,090	101,614	0.63	$10,800	1997	$21,600	0.79	$14,936	2.8	$42,526	$15,231	448 gpm, 150 ft head	WP222	17.83
P-223	1	0	Lime Unloading Blower	STRM0227	547	337	0.62	$47,600	1998	$47,600	0.5	$37,340	1.4	$52,898	$37,785	3341 cfm, 6 psi, 10,024 lb/hr	WP223	4.10
P-224	1	1	Hydrolysis Feed Pump	STRM0250	160,000	167,795	1.05	$64,934	1999	$129,868	0.6	$133,628	1.2	$160,354	$133,628	740 gpm, 240 ft head	WP224	119.31
P-225	1	1	ISEP Elution Pump	STRM0243	52,731	18,005	0.34	$7,900	1997	$15,800	0.79	$6,761	2.8	$19,249	$6,894	104 gpm, 150 ft head	WP225	3.92
P-226	1	1	ISEP Reload Pump	STRM0246	164,080	100,802	0.61	$8,700	1997	$17,400	0.79	$11,841	2.8	$33,714	$12,075	445 gpm, 150 ft head	WP226	17.92
P-227	1	1	ISEP Hydrolyzate Feed Pump	STRM0221	160,290	98,157	0.61	$10,700	1997	$21,400	0.79	$14,526	2.8	$41,359	$14,814	432 gpm, 150 ft head	WP227	16.81
P-239	1	1	Reacidified Liquor Pump	STRM0239	167,280	102,752	0.61	$10,800	1997	$21,600	0.79	$14,698	2.8	$41,847	$14,988	450 gpm, 100 ft head	WP239	12.09
S-202	3	0	Pre-IX Belt Filter Press	SOLD0220	57,000	57,000	1.00	$200,000	1998	$600,000	0.39	$600,000	1.4	$850,010	$607,150	Use 3 units for 45% of the flow as recommended by the vendor	WS202	19.69
S-221	1	0	ISEP	STRM0240	210,005	98,157	0.47	$2,058,000	1997	$2,058,000	0.33	$1,601,194	1.2	$1,959,422	$1,632,851	10 chambers (39" dia. X 84" high), 4" dia. Valve - Weak Base Resin	WS221	2.98
S-222	1	0	Hydroclone & Rotary Drum Filter	STRM0229	5,195	1,137	0.22	$165,000	1998	$165,000	0.39	$91,224	1.4	$129,235	$92,311	Hydrocyclone and Vacuum Filter for 453 gpm	WS222	11.93
S-227	1	0	LimeDust Vent Baghouse	STRM0227	548	337	0.61	$32,200	1997	$32,200	1	$19,778	1.5	$30,254	$20,169	3750 cfm, 625 sf, 6 cfm/sf		
T-201	1	0	Sulfuric Acid Storage	STRM0710	1,647	860	0.52	$5,760	1996	$5,760	0.71	$3,633	1.7	$6,283	$3,751	2000 gal., 24 hr. residence time, 90% wv, 5.5ft diam. X 11ft		
T-203	1	0	Blowdown Tank	STRM0217	270,300	121,514	0.45	$64,100	1997	$64,100	0.93	$30,475	1.7	$52,061	$31,078	7000 gal., 11' dia x 30' high, 10 min. res. time, 75% wv, 15 psig		
T-209	1	0	Overliming Tank	STRM0228	167,050	102,608	0.61	$71,000	1997	$71,000	0.71	$50,232	1.8	$90,186	$51,225	29850 gal., 16' dia. X 32' high, 1 hr. res. time, 90% wv, 15 psig		
T-220	1	0	Lime Storage Bin	STRM0227	548	548	1.00	$69,200	1997	$69,200	0.46	$69,200	1.8	$124,243	$70,568	4455 cf, 14' dia x 25' high, 1.5x rail car vol., atmospheric, 15 day storage max		
T-224	1	0	Reacidification Tank	STRM0239	102,752	102,752	1.00	$111,889	1999	$111,889	0.51	$111,889	1.8	$196,992	$111,889	120,000 gal., 28' dia x 28' high, 4 hr. res. time, 90% wv, atmospheric		
T-232	1	0	Slurrying Tank	STRM0250	358,810	167,795	0.47	$44,800	1997	$44,800	0.71	$26,117	1.8	$46,890	$26,633	11300 gal., 13' dia. X 25' high, 15 min. res. time, 90% wv		
0	0	0	0	0	0	0	0.00	$0	1999	$0	0	$0	-	$0	$0	0		
										weighted averages:	0.70		1.48					676.27
A200									Subtotal	$16,527,758		$9,999,337		$14,955,166	$10,128,493			
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.5	$15,025,380	$70,213	is installed cost savings		

A-300	8	0	Fermentor Agitators	GALLONS	962,651	750,000	0.78	$19,676	1996	$157,408	0.51	$138,592	1.2	$175,799	$143,110	Side Mounted, 2 per vessel, 60 hp each, 0.15 hp/1000 gal	WT300	201.34
A-301	1	0	Seed Hold Tank Agitator	STRM0304	41,777	17,529	0.42	$12,551	1996	$12,551	0.51	$8,060	1.2	$10,223	$8,322	Top Mounted, 1800 rpm, 10 hp, 0.1 hp/1000 gal	WT301	5.59
A-304	2	0	4th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$11,700	1997	$23,400	0.51	$15,026	1.2	$18,824	$15,323	Top Mounted, 1800 rpm, 3 hp, 0.3 hp/1000 gal	WT304	3.36
A-305	2	0	5th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$10,340	1996	$20,680	0.51	$13,280	1.2	$16,845	$13,713	Top Mounted, 1800 rpm, 9 hp, 0.1 hp/1000 gal	WT305	10.07
A-306	1	0	Beer Well Agitator	STRM0502	381,700	173,737	0.46	$10,100	1997	$10,100	0.51	$6,761	1.2	$8,469	$6,894	Top Mounted, 1800 rpm, 2 hp, 0.3 hp/1000 gal	WT306	1.12
F-300	4	0	Fermentors	GALLONS	750,000	750,000	1.00	$326,203	1999	$1,304,812	0.71	$1,304,812	1.8	$2,297,260	$1,304,812	750,000 gal. each, 2 day residence total, 90% wv, API, atmospheric, 50' f x 51'		
F-301	2	0	1st Fermentation Seed Fermentor	None		0	0.45	$14,700	1997	$29,400	0.93	$13,991	2.8	$39,948	$14,267	9 gal, jacketed, agitated, 1' dia., 1.5' high, 15 psig		
F-302	2	0	2nd Fermentation Seed Fermentor	None		0	0.45	$32,600	1997	$65,200	0.93	$31,027	2.8	$88,592	$31,640	90 gal., jacketed, agitated, 2' 3" dia., 3' high, 2.5 psig		
F-303	2	0	3rd Fermentation Seed Fermentor	None		0	0.45	$81,100	1997	$162,200	0.93	$77,186	2.8	$220,394	$78,712	900 gal., jacketed, agitated, 5' dia, 6.5' high, 2.5 psig		
F-304	2	0	4th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$39,500	1997	$79,000	0.93	$35,225	1.7	$60,174	$35,921	9000 gal., 9' dia x 19' high, atmospheric		
F-305	2	0	5th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$147,245	1998	$294,490	0.51	$189,107	1.8	$336,910	$191,360	90000 gal., API, atmospheric 25'f x 25'		
H-300	4	1	Fermentation Cooler	QHX300EA	67,820	25,053	0.37	$4,000	1997	$20,000	0.78	$9,198	2.2	$20,438	$9,380	4 exchangers at 221 sf, U=300 BTU/hr sf F LMTD = 22.9°F plate and frame		
H-301	1	0	Fermentation Seed Hydrolyzate Cooler	AREA0301	773	318	0.41	$15,539	1998	$15,539	0.78	$7,778	2.2	$17,151	$7,871	348 sf, 300 BTU/hr sf F		
H-302	1	0	Fermentation Pre-Cooler	AREA0302	3,765	828	0.22	$25,409	1998	$25,409	0.78	$7,797	2.2	$17,193	$7,890	828 sf total, plate and frame		
H-304	1	0	4TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$3,300	1997	$3,300	0.83	$1,580	1.2	$1,934	$1,611	12 sf, 1" sch 40 pipe, 105 BTU/hr sf F		
H-305	1	0	5TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$18,800	1997	$18,800	0.98	$7,881	1.2	$9,644	$8,037	138 sf, 2" sch 40 pipe, 92 BTU/hr sf F		
P-300	4	1	Fermentation Recirc./Transfer Pump	QHX300EA	67,737	55,505	0.82	$8,000	1997	$40,000	0.79	$34,177	2.8	$97,307	$34,852	844 gpm @ 150 ft sized based on heating rate	WP300	104.49
P-301	1	1	Fermentation Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$22,194	1998	$44,388	0.7	$24,168	1.4	$34,238	$24,456	280 gpm @ 150 ft head	WP301	5.95
P-302	2	0	Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$54,088	1998	$108,176	0.7	$58,898	1.4	$83,440	$59,600	504 gpm total, 252 gpm each, 100 ft head	WP302	7.14
P-306	1	1	Beer Transfer Pump	STRM0502	381,701	173,737	0.46	$17,300	1997	$34,600	0.79	$18,579	2.8	$52,899	$18,947	790 gpm each, 171 ft head	WP306	34.47
T-301	1	0	Fermentation Seed Hold Tank	STRM0304	41,777	17,529	0.42	$161,593	1998	$161,593	0.51	$103,767	1.8	$184,870	$105,003	105000 gal., API atmospheric		
T-306	1	0	Beer Well	STRM0502	129,000	183,467	1.42	$111,889	1999	$111,889	0.51	$133,906	1.8	$235,756	$133,906	192,518 gal., 32' dia x 32' high, 4 hr. res. time, 95% wv, atmospheric		
										weighted averages:	**0.68**		**1.79**					373.53
A300									**Subtotal**	**$2,742,935**		**$2,240,795**		**$4,028,307**	**$2,255,629**			
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.3	$8,218,509	$4,190,202	is installed cost savings		
A-307	8	0	Enzymatic Hydrolysis Tank Agitators	STRM0302B	157,136	157,136	1.00	$19,676	1996	$157,408	0.51	$157,408	1.2	$199,666	$162,539	two side mounted 75 hp agitators / tank, 0.4hp/1000 gal.	WT307	251.67
H-307	12	0	Enzymatic Hydrolysis Tank Heater	STRM0302B	157,136	157,136	1.00	$15,000	1999	$180,000	0.78	$180,000	2.2	$392,214	$180,000	65 ft2 double pipe		
H-308	1	0	Pre-hydrolyzate cooler	STRM0302	145,536	145,536	1.00	$25,000	1999	$25,000	0.78	$25,000	2.2	$54,474	$25,000	481 ft2, parallel double pipe		
P-308	8	1	Hydrolyzer Bottoms Pump	STRM0302B	157,136	157,136	1.00	$121,690	1999	$1,095,210	0.6	$1,095,210	1.2	$1,314,252	$1,095,210	3000 GPM each Disc flow pumps, 245ft head	WP308	1,744.94
T-307	4	0	Enzymatic Hydrolysis Tank	STRM0302B	750,000	375,000	0.50	$326,203	1999	$1,304,812	0.6	$860,855	2.0	$1,753,728	$860,855	375,000 gallons, 24 hour residence time, 2 side mounted agitators cone bottom, concrete base, bottom outlet through the concrete, 30o cone		
0	0	0	0	0	0	0	0.00	$0	1999	$0	0	$0	-	$0	$0	0		
										weighted averages:	**0.61**		**1.60**					1,996.61
Area 307									**Subtotal**	**$2,762,430**		**$2,318,473**		**$3,714,334**	**$2,323,604**			
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			-	$0	($3,714,334)	is installed cost savings		
															$475,868	***sizes not updated to reflect 25% increase, but costs and energy consumed are upd***		
A-400	11		Cellulase Fermentor Agitators	GALLONS	150,000	117,779.84	0.79	$ 200,000	1999	$2,200,000	0.51	$ 1,944,743	1.2	$2,388,960	$1,944,743	125 hp / agitator -- 1 agitator/vessel	WT400	745.70
F-400	11		Cellulase Fermentors	GALLONS	88,335	117,779.84	1.33	$ 179,952	1998	$1,979,472	0.71	$ 2,428,040	1.8	$4,325,765	$2,456,975	88335 gal, 2.5 psig, cooling coils in tank costed as H400, 40 ft. height, 20 ft. diameter		
F-401	3		1st Cellulase Seed Fermentor	STRM0433	2,790	1,242.43	0.45	$ 22,500	1997	$67,500	0.93	$ 31,810	2.0	$64,878	$32,439	11 gal / 15 psig / Jacketed / Agitator		
F-402	3		2nd Cellulase Seed Fermentor	STRM0433	2,790	1,242.43	0.45	$ 54,100	1997	$162,300	0.93	$ 76,486	2.0	$155,996	$77,998	221 gal / 15 psig / Jacketed /Agitator	WT402	159.77
F-403	3		3rd Cellulase Seed Fermentor	STRM0433	2,790	1,242.43	0.45	$ 282,100	1997	$846,300	0.93	$ 398,829	2.0	$813,429	$406,715	4417 gal / 15 psig / Jacketed./Agitator		
H-400	11		Cellulase Fermentation Cooler	QHX400EA	236,668	117,779.84	0.50	$ 34,400	1997	$378,400	0.78	$ 219,562	2.2	$487,878	$223,903	Immersible Coil 205 ft2 each		
M-401	5	1	Fermentor Air Compressor Package	STRM0440	80,455	107,273.33	1.33	$ 229,000	1999	$1,374,000	0.34	$ 1,515,186	1.3	$1,969,742	$1,515,186	7946 scfm each, 50 psig outlet, 1277 hp each, includes starter	WM401	6,810.67
P-400	1	1	Cellulase Transfer Pump	STRM0420	40,543	15,467.03	0.38	$ 9,300	1997	$18,600	0.79	$ 8,687	2.8	$24,735	$8,859	58 GPM / 100 ft. head	WP400	2.10
P-401	1	1	Cellulase Seed Pump	STRM0433	2,790	1,242.43	0.45	$ 12,105	1998	$24,210	0.70	$ 13,742	1.2	$16,687	$13,906	24 gpm / 1 hp	WP401	0.37
P-405	1	1	Media Pump	STRM0416	586	266.85	0.46	$ 8,300	1997	$16,600	0.79	$ 8,917	2.8	$25,388	$9,093	21 Gpm/100 Ft Head	WP405	0.12
P-420	1	1	Anti-foam Pump	STRM0417	227	104.85	0.46	$ 5,500	1997	$11,000	0.79	$ 5,976	2.8	$17,013	$6,094	4 gpm / 75 ft head	WP420	0.02
T-405	1		Media-Prep Tank	STRM0416	586	266.85	0.46	$ 64,600	1997	$64,600	0.71	$ 36,955	1.7	$63,130	$37,685	2083 Gal / 1.17 hp Agitator	WT402	0.87
T-420	1		Anti-foam Tank	STRM0417	227	104.85	0.46	$ 402	1998	$402	0.71	$ 232	1.7	$394	$235	67 gal, 3 hr. residence time		
												area install factor	1.5					7,719.61
A400									**Subtotal**	**$7,143,384**		**$6,689,166**		**$10,353,995**	**$6,733,832**			
														$10,353,995				

D-501	1	0	Beer Column	DIAMD501	4	2	0.56	$636,976	1996	$636,976	0.78	$402,792	2.1	$873,434	$415,921	7'6" DIA, 32 ACTUAL TRAYS, NUTTER V-GRID TRAYS		
D-502	1	0	Rectification Column	S510S521	56,477	26,744	0.47	$525,800	1996	$525,800	0.78	$293,491	2.1	$636,421	$303,058	8' dia.(rect)., 4' dia. (strip) x 18" T.S., 60 act. Trays, 60% eff., Nutter V-Grid trays		
E-501	1	0	1st Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,676	1996	$435,676	0.68	$435,676	2.1	$944,742	$449,877	22278 sf each., 135 BTU/hr sf F		
E-502	1	0	2nd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.1	$944,685	$449,850	22278 sf., 170 BTU/hr sf F		
E-503	1	0	3rd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.1	$944,685	$449,850	22278 sf each., 170 BTU/hr sf F		
H-501	1	0	Beer Column Reboiler	QRFD0501	-7,863,670	-3,723,722	0.474	$158,374	1996	$158,374	0.68	$95,263	2.2	$214,340	$98,368	Fixed TS, 6602 sf, 31" dia., 20' long, 178 BTU/hr sf F		
H-502	1	0	Rectification Column Reboiler	QRFD0502	-987,427	-467,581	0.474	$29,600	1997	$29,600	0.68	$17,805	2.2	$39,563	$18,157	Thermosyphon, 512 sf, 15" dia., 20' long, 130 BTU/hr sf F		
H-504	1	0	Beer Column Condenser	QCND0501	277,820	131,557	0.474	$29,544	1996	$29,544	0.68	$17,771	2.2	$39,984	$18,350	Floating Head, 418 sf, 15" dia., 22' long, 92 BTU/hr sf F		
H-505	1	0	Rectification Column Condenser	QCND0502	4,905,410	2,322,883	0.474	$86,174	1996	$86,174	0.68	$51,834	2.2	$116,626	$53,524	Fixed TS, 1969 sf, 29" dia, 20' long, 157 BTU/hr sf F		
H-512	1	1	Beer Column Feed Interchange	AREA0512	909	430	0.474	$19,040	1996	$38,080	0.68	$22,905	2.2	$51,537	$23,652	431 sf, 200 BTU/hr sf F		
H-517	1	1	Evaporator Condenser	QHET0517	6,764,222	3,203,095	0.47	$121,576	1996	$243,152	0.68	$146,257	2.2	$329,077	$151,024	Fixed TS, 3906 sf, 29" dia., 20' long, 220 BTU/hr sf F		-
M-503	1	0	Molecular Sieve (9 pieces)	STRM0515	20,491	9,703	0.47	$2,700,000	1998	$2,700,000	0.7	$1,599,964	1.0	$1,619,030	$1,619,030	Superheater, twin mole sieve columns, product cooler, condenser, pumps, vacuum source.	WM503	55.00
P-501	1	1	Beer Column Bottoms Pump	P501FLOW	5,053	2,200	0.44	$42,300	1997	$84,600	0.79	$43,861	2.8	$124,881	$44,728	2200 gpm, 150 ft head	WP501	84.65
P-503	1	1	Beer Column Reflux Pump	QCND0501	277,820	131,557	0.47	$1,357	1998	$2,714	0.79	$1,504	2.8	$4,248	$1,522	6 gpm, 140 ft head	WP503	0.22
P-504	1	1	Rectification Column Bottoms Pump	STRM0516	31,507	15,530	0.49	$4,916	1998	$9,832	0.79	$5,622	2.8	$15,884	$5,689	76 gpm, 158 ft head	WP504	2.80
P-505	1	1	Rectification Column Reflux Pump	QCND0502	4,906,301	2,323,304	0.47	$4,782	1998	$9,564	0.79	$5,299	2.8	$14,970	$5,362	207 gpm, 110 ft head	WP505	5.14
P-511	2	1	1st Effect Pump	STRM0525	278,645	133,617	0.48	$19,700	1997	$59,100	0.79	$33,069	2.8	$94,155	$33,723	1137 gpm each, 110 ft head	WP511	67.89
P-512	1	1	2nd Effect Pump	STRM0528	91,111	45,390	0.50	$13,900	1997	$27,800	0.79	$16,032	2.8	$45,646	$16,349	599 gpm, 110 ft head	WP512	17.37
P-513	2	1	3rd Effect Pump	STRM0531	48,001	23,814	0.50	$8,000	1997	$24,000	0.79	$13,795	2.8	$39,276	$14,068	196 gpm each, 110 ft head	WP513	12.54
P-514	1	1	Evaporator Condensate Pump	STRM534A	140,220	69,285	0.49	$12,300	1997	$24,600	0.79	$14,095	2.8	$40,131	$14,374	293 gpm, 125 ft head	WP514	9.20
P-515	1	1	Scrubber Bottoms Pump	STRM0551	15,377	7,427	0.48	$2,793	1998	$5,586	0.79	$3,143	2.8	$8,881	$3,181	31 gpm, 104 ft head	WP515	0.84
P-517	1	1	Kill Tank Bottoms Pump	STRM0518	5,053	660	0.13	$42,300	1997	$84,600	0.79	$16,944	2.8	$48,242	$17,279	660gpm, 72 ft head	WP517	12.19
T-503	1	0	Beer Column Reflux Drum	QCND0501	277,820	131,557	0.47	$11,900	1997	$11,900	0.93	$5,938	1.7	$10,144	$6,055	164 gal, 15 min res. Time, 50% wv, 2'6" dia., 5' long, 25 psig		
T-505	1	0	Rectification Column Reflux Drum	QCND0502	4,906,301	2,323,304	0.47	$45,600	1997	$45,600	0.72	$26,621	1.7	$45,476	$27,147	6225 gal, 15 min res time, 50% wv, 7' dia, 22' long, 25 psig		
T-512	1	0	Vent Scrubber	STRM0523	18,523	9,788	0.53	$99,000	1998	$99,000	0.78	$60,197	1.7	$102,043	$60,915	5' dia x 25' high, 4 stages, plastic Jaeger Tri-Packing		
T-513	1	0	Kill Tank	STRM0518	149,897	149,897	1.00	$99,920	1999	$99,920	0.78	$99,920	1.7	$167,384	$99,920	18 psig, 30 min. res. time		
										weighted averages:	0.72		1.7					267.85
A500									**Subtotal**	$6,343,492		$4,301,097		**$7,515,486**	$4,400,972			
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.7	$6,765,614	-$749,872	is installed cost savings		

C-601	1	0	Lignin conveyor	STRM0601B	225,140	225,140	1.00	$31,700	1997	$31,700	0.60	$31,700	1.5	$49,832	$32,327	14" dia. 100' long	WC109	21.50
M-613	1	0	Syrup Sprayer	STRM0531	22,372	22,372	1.00	$1,000	1999	$1,000	0.3	$1,000	1.2	$1,200	$1,000	100 GPM syrup sprayer		
M-614	1	0	Lignin Loadout	STRM0601A	63,778	0	0.00	$41,200	1999	$41,200	0.3	$0	1.0	$0	$0	245 GPM @ 20.6% insoluble solids		
M-615	1	0	Equalization Basin	STRM0830	98,267	102,204	1.04	$350,000	1999	$350,000	0.79	$361,031	1.0	$361,031	$361,031	no less than 500,000 gal., above-ground bolted tank with cover, including foundations, pumps and controls	WM615	1,077.21
M-616	1	0	Anaerobic Digestion System	STRM0830	98,267	102,204	1.04	$3,200,000	1999	$3,200,000	0.79	$3,300,852	1.0	$3,300,852	$3,300,852	500,000 gal., includes site work, foundations, reactors and ancillary equipment		
M-617	1	0	Aerobic Digestion System	STRM0830	98,267	102,204	1.04	$4,300,000	1999	$4,300,000	0.79	$4,435,520	1.0	$4,435,520	$4,435,520	four-350,000 gal. Sequencing Batch Reactors, 48,000 lbs/day of O2 transfer capability, de-nitrification facilities, aeration and mixing requires approximately 1,400 horsepower		
M-618	1	0	Pressure Sand Filters	STRM0830	98,267	102,204	1.04	$280,000	1999	$280,000	0.79	$288,825	1.0	$288,825	$288,825	400 ft2 of filtration surface area, includes the engineering and legal cost to acquire an NPDES permit		
P-630	1	1	Recycle Water Pump	STRM0602	179,446	84,120	0.47	$10,600	1997	$21,200	0.79	$11,652	2.8	$33,175	$11,882	370 gpm, 150ft head	WP630	14.75
S-601	2	0	Beer Column Bottoms Centrifuge	CENTFLOW	404	300	0.74	$659,550	1998	$1,319,100	0.6	$1,103,371	1.2	$1,339,824	$1,116,520	requires 540gpm duty, 2 @ 300 gpm and 410 hp each	WS601	489.18
T-630	1	0	Recycled Water Tank	STRM0602	179,446	84,120	0.47	$14,515	1998	$14,515	0.745	$8,254	1.7	$13,992	$8,353	7410 gal, 20 min. res., 2.5 psig, 9.5ft diam. x 14.25ft		
										weighted averages:	**0.7609184**		1.0					1,602.64
A600									**Subtotal**	**$9,558,715**		**$9,542,206**		**$9,824,251**	**$9,556,310**			
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.3	$5,167,342	($4,656,910)	is installed cost savings		

																sizes not updated to reflect 25% increase, but costs and energy consumed are upd		
P-703	1	1	Sulfuric Acid Pump	STRM0710	1,647	1,912	1.16	$8,000	1997	$16,000	0.79	$18,001	2.8	$51,252	$18,357	215 gpm, 150ft head	WP703	[redacted]
P-707	1	1	Antifoam Store Pump	STRM0417	227	105	0.46	$5,700	1997	$11,400	0.79	$6,193	2.8	$17,633	$6,315	0.5 gpm, 92 ft head	WP707	0.01
P-720	1	1	CSL Pump	STRM0735	2,039	859	0.42	$8,800	1997	$17,600	0.79	$8,889	2.8	$25,309	$9,065	182 gpm, 150ft head	WP720	0.15
T-703	1		Sulfuric Acid Storage Tank	STRM0710	1,647	1,912	1.16	$42,500	1997	$42,500	0.51	$45,860	1.8	$82,338	$46,767	20,000 gal, 240 hr supply, 90% wv, 12ft diam. x 24 ft, atmospheric		
T-707	1		Antifoam Storage Tank	STRM0417	227	303	1.33	$14,400	1997	$14,400	0.71	$17,663	1.7	$30,174	$18,012	12,000 gal, 27 day supply, 10.5ft diam. X 18.5ft		
T-720	1		CSL Storage Tank	STRM0735	2,039	859	0.42	$88,100	1997	$88,100	0.79	$44,495	1.7	$76,011	$45,375	30160 gal, 90% wv, 120 supply, 14.3ft diam. X 25 ft		
											area install factor		2.0					0.28
A700									**Subtotal**	**$190,000**		**$141,101**		**$282,716**	**$143,891**			
					2000tpd x .45 (current year cost with area weighted-average scale exponent applied)								1.5	$2,171,166	$1,888,451	is installed cost savings		
M-803	1	0	Boiler with Superheater	STRM0815 + 216	200,000	200,000	1.00	$1,590,000	1999	$1,590,000	0.7	$1,590,000	1.3	$2,067,000	$1,590,000	200,000 #/hr running @ 171,488 #/hr; with 40,000 #/hr 160o superheat; 132,000#/hr 390o sat. @ 205 psig	WM803	75.60
M-820	1	0	Hot process water softener system	STRM0811B	229,386	45,003	0.20	$1,383,300	1999	$1,383,300	0.6	$520,623	1.2	$624,748	$520,623	200 gpm		
M-830	1	0	Hydrazine Addition Pkg.	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM830	10.00
M-832	1	0	Ammonia Addition Pkg	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM832	10.00
M-834	1	0	Phosphate Addition Pkg.	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM834	10.00
P-804	2	1	Condensate Pump	STRM811A	249,633	38,798	0.16	$7,100	1997	$21,300	0.79	$4,894	4.6	$22,958	$4,991	130 gpm, 150' head	WP804	9.21
P-824	2	1	Deaerator Feed Pump	STRM811A	196,000	38,798	0.20	$9,500	1997	$28,500	0.79	$7,927	8.3	$67,097	$8,084	180 gpm, 115' head	WP824	4.89
P-826	4	1	BFW Pump	STRM0813	207,310	80,536	0.39	$52,501	1998	$262,505	0.79	$124,377	1.4	$176,203	$125,859	310 gpm, 2740' head	WP826	400.99
P-828	1	1	Blowdown Pump	STRM0821	6,600	2,699	0.41	$5,100	1997	$10,200	0.79	$5,032	6.4	$32,842	$5,132	12 gpm, 150' head	WP828	0.42
P-830	1	1	Hydrazine Transfer Pump	STRM813A	229,386	80,536	0.35	$5,500	1997	$11,000	0.79	$4,811	6.4	$31,402	$4,907	3 gpm, 75' head	WP830	0.05
T-804	1	0	Condensate Collection Tank	STRM811A	229,386	38,798	0.17	$7,100	1997	$7,100	0.71	$2,011	3.3	$6,766	$2,050	200 gal, 1.5 min. res. time		
T-824	1	0	Condensate Surge Drum	STRM811A	150,000	38,798	0.26	$49,600	1997	$49,600	0.72	$18,734	5.0	$95,523	$19,105	2100 gal., 6' diam. X 10', 15 psig, res. time 11 min.		
T-826	1	0	Deaerator	STRM0813	267,000	80,536	0.30	$165,000	1998	$165,000	0.72	$69,616	6.5	$457,896	$70,446	3030 gal., 15 psig, 10 min. res.		
T-828	1	0	Blowdown Flash Drum	STRM0821	6,550	2,699	0.41	$9,200	1997	$9,200	0.72	$4,859	7.3	$36,168	$4,955	210 gal., 2.5' diam. X 6', 50 psig 17 min. res.		
T-830	1	0	Hydrazine Drum	STRM813A	229,386	80,536	0.35	$12,400	1997	$12,400	0.93	$4,685	7.0	$33,440	$4,777	138 gal, 3.75' x 1.25' diam., 10 psig		
										weighted averages:	**0.6704429**		1.5					521.16
A800									**Subtotal**	**$3,607,105**		**$2,387,986**		**$3,684,612**	**$2,393,497**			
					2000tpd x .45 (current year cost with area weighted-average scale exponent applied)								1.1	$23,046,972	$19,362,360	is installed cost savings		
M-902	1	0	Cooling Tower System	QCWCAPIT	41,100,000	12,955,985	0.32	$1,659,000	1998	$1,659,000	0.78	$674,181	1.2	$818,659	$682,216	40,000 gpm, 185.4MM BTU/hr	WM902	298.85
M-904	1	0	Plant Air Compressor	STRM0101	159,950	159,950	1.00	$60,100	1997	$60,100	0.34	$60,100	1.3	$79,675	$61,288	450 cfm, 125 psig outlet	WM904	186.40
M-908	1	0	Chilled Water Package	QCHLWCAP	5,040,000	2,268,000	0.45	$380,000	1997	$380,000	0.8	$200,610	1.2	$245,492	$204,577	1000 ton, 600kW	WM908	600.00
M-910	1	0	CIP System	STRM0914	63	28	0.45	$95,000	1995	$95,000	0.6	$58,837	1.2	$73,021	$60,851	designed by Delta-T, (est 0.2 kW)	WM910	0.20
P-902	1	1	Cooling Water Pumps	STRM0940	18,290,000	5,553,791	0.30	$332,300	1997	$664,600	0.79	$259,201	2.8	$737,993	$264,326	12300 gpm, 70ft head		
P-912	1	1	Make-up Water Pump	STRM0904	244,160	82,445	0.34	$10,800	1997	$21,600	0.79	$9,161	2.8	$26,084	$9,343	370 gpm, 75ft head	WP912	7.32
P-914	1	1	Process Water Circulating Pump	STRM0905	352,710	111,503	0.32	$11,100	1997	$22,200	0.79	$8,938	2.8	$25,449	$9,115	745 gpm, 75ft head	WP914	14.78
S-904	1	1	Instrument Air Dryer	STRM0101	159,950	71,977	0.45	$15,498	1999	$30,996	0.6	$19,197	1.3	$24,956	$19,197	134 scfm air dryer, -40F Dewpoint	WS601	4.91
T-904	1	0	Plant Air Receiver	STRM0101	159,950	53,316	0.33	$13,000	1997	$13,000	0.72	$5,894	1.7	$10,069	$6,011	300 gal., 200 psig		
T-914	1	0	Process Water Tank	STRM0905	352,710	111,503	0.32	$195,500	1997	$195,500	0.51	$108,663	1.8	$195,095	$110,811	234360 gal, 8hr res. time		
										weighted averages:	**0.751991**		**1.57**			400 gpm well pump, 500ft head	53.16	1,165.62
Area 900									**Subtotal**	**$3,141,996**		**$1,404,783**		**$2,236,491**	**$1,427,733**		**Total kW**	**14,823**
					2000tpd x .45 (current year cost with area weighted-average scale exponent applied)								1.3	$5,278,320	$3,041,829	is installed cost savings		

PLANT TOTAL:	**$57,333,793**	**$44,443,850**	**$62,741,793**
45% NREL TOTAL:			**$79,208,934**
SAVINGS: (do to much cheaper boiler and effect of separation of hydrolysis and fermentation)			**$16,467,141**
			20.79%

TUDY MODEL WITH REFRENCE MODEL CELLULASE PRODUCTION: CELLULASE PRODUCTION COST (as prorated based on fraction of liquor requ

CAPITAL INVESTMENT ASSUMPTIONS

1) Total capital investment

Civil Structural		$ 500,000	estimated
Area 100		$ 308,790	
Area 200		$ 751,332	
Area 300		$ -	
Area 307		$ -	
Area 400		$ 10,353,995	
Area 500		$ -	
Area 600		$ 430,086	
Area 700		$ 176,840	
Area 800		$ 64,163	
Area 900		$ 38,945	estimated
Fixed Capital		$ 12,624,151	
INDIRECTS Prorateable	3.5%	$441,845	
Process Development	2.0%	$252,483	
Field Expense	8.0%	$1,009,932	
Home Office Constr. Fee	12.0%	$1,514,898	
Contingency	10.0%	$1,262,415	
Start-up, Permits, Fees	3.0%	$378,725	
Working Capital per estimate		$123,114	1 mos Raw matls. + O&M
Total Plant Cost		$17,607,563	
FEDERAL & STATE GRANTS	10%	($1,760,756)	
Net Capital Investment		**$15,846,807**	

OPERATING COST ASSUMPTIONS 8,400 hr/yr

		Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Electricity		7,910	Kw-hr	$0.035	$277	$2,325,687
well water		0	kg	$0.000	$0	$0
Wastewater		2,977	kg	$0.00026	$1	$6,605
gypsum waste disposal	($33/ston)	57	kg	$0.0364	$2	$17,449
Total Utilities					$280	$2,349,741

Raw Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Corn Stover DRY (stm 101 less water)	1,884	kg	$0.016	$30.01	$252,113
Sulfuric Acid (stm 710)	43	kg	$0.100	$4.33	$36,403
Calcium Oxide (Lime stm 227)	17	kg	$0.293	$4.96	$41,650
Ammonia (stm 717)	73	kg	$0.162	$11.86	$99,586
Corn Steep Liquor (stm 735)	200	kg	$0.051	$10.21	$85,801
Nutrients	80	kg	$0.291	$23.31	$195,794
Cellulase Complex	0	kg	$3.000	$0.00	$0
Natural Gasoline (stm 701)	0	kg	$0.155	$0.00	$0
Diesel/Gasoline	4	kg	$0.155	$0.62	$5,198
WWT Chemicals	0.2	kg	$2.237	$0.52	$4,404
CW Chemicals	0.3	kg	$1.428	$0.42	$3,566
BFW Chemicals	1.3	kg	$0.226	$0.29	$2,435
Boiler Fuel (stm 813)	3	Mbtu	$2.500	$8.29	$69,637
Total Raw Materials				$95	$796,588

$per lb. calcs.

Processing Material Costs	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Antifoam (Corn Oil)	105	kg	$0.304	$32	$267,948
Total Processing Materials				$32	$267,948

Operations and Maintenance Costs - DRY HANDLING (area 100)	each/day		wage	hr/day each	Total Cost /yr.
Supervisors	0.025	$	20.00	12	$2,200
Operators	0.100	$	16.00	12	$7,041
Laborers	0.402	$	16.00	12	$28,166
Maintenance	0.100	$	16.00	12	$7,041
Operations and Maintenance Costs - HYDROLYSIS/FERMENTATION (area 200, 300, 400, 500, 600)					
Supervisors	0.0	$	20.00	12	$2,200
Operators	0.2	$	16.00	8	$7,041
Laborers	0.1	$	16.00	8	$2,347
Technicians (Includes Lab.)	0.2	$	16.00	8	$7,041
Maintenance	0.2	$	16.00	8	$7,041
Operations and Maintenance Costs - Utilities (area 700, 800, 900)					
Supervisors	0.0	$	20.00	12	$1,100
Operators	0.2	$	16.00	8	$3,521
Laborers	0.1	$	16.00	8	$1,174
Technicians	0.1	$	16.00	8	$1,174
Maintenance	0.1	$	16.00	8	$2,347
Total Operations and maintenance labor costs					$79,437

Other Operations and Maintenance Costs			
Payroll Overhead	35% of operating labor	$	27,803
Maintenance Costs	2% of plant cost	$	252,483
Operating Supplies	0.25% of plant cost	$	31,560
Environmental	0.50% of plant cost	$	63,121
Local Taxes	1% of plant cost	$	126,242
Insurance	0.50% of plant cost	$	63,121
Overhead Costs	40% of labor, supervision,maint cost	$	31,775
Administrative Costs	1% of annual sales (less tax credits)	$	5,239
Distribution and Sales	0.5% of annual sales (less tax credits)	$	-
Total O&M Costs			$680,780

Operating Expenses:		
Utilities		2,349,741
Raw Materials		796,588
Processing Materials		267,948
Operation & Maintenance		680,780
Property Tax @	0.50% Book Value	79,234
Depreciation		1,056,454
Debt retierment		1,418,471
Total Operating Expense ($/yr)		**$6,649,217**

IIGH PLAINS YORK CELLULASE PRODUCTION WITH PURVISION TECHNOLOGY (as prorated based on fraction of liquor required)

CAPITAL INVESTMENT ASSUMPTIONS

1) Total capital investment

Civil Structural		$	500,000	estimated
Area 100		$	204,724	
Area 200		$	381,629	
Area 300		$	-	
Area 307		$	-	
Area 400		$	5,692,516	
Area 500		$	-	
Area 600		$	313,763	
Area 700		$	122,171	
Area 800		$	31,259	
Area 900		$	18,646	estimated
Fixed Capital		$	7,264,708	
INDIRECTS Prorateable	3.5%		$254,265	
Process Development	2.0%		$145,294	
Field Expense	8.0%		$581,177	
Home Office Constr. Fee	12.0%		$871,765	
Contingency	10.0%		$726,471	
Start-up, Permits, Fees	3.0%		$217,941	
Working Capital per estimate			$84,188	1 mos Raw matls. + O&M
Total Plant Cost			$10,145,809	
FEDERAL & STATE GRANTS	10%		($1,014,581)	
Net Capital Investment			**$9,131,228**	

OPERATING COST ASSUMPTIONS **8,400** hr/yr

		Amount/hr	*Units*	*$/unit*	*Cost /hr.*	**Total Cost /yr**
Electricity		5,918	Kw-hr	$0.035	$207	$1,739,954
well water		0	kg	$0.000	$0	$0
Wastewater		2,232	kg	$0.00026	$1	$4,954
gypsum waste disposal	($33/ston)	43	kg	$0.0364	$2	$13,087
Total Utilities					$209	$1,757,995

Raw Material Costs

	Amount/hr	*Units*	*$/unit*	*Cost /hr.*	**Total Cost /yr**
Corn Stover DRY (stm 101 less water)	1,413	kg	$0.016	$22.51	$189,083
Sulfuric Acid (stm 710)	32	kg	$0.100	$3.25	$27,302
Calcium Oxide (Lime stm 227)	13	kg	$0.293	$3.72	$31,237
Ammonia (stm 717)	57	kg	$0.162	$9.30	$78,093
Corn Steep Liquor (stm 735)	151	kg	$0.051	$7.69	$64,629
Nutrients	60	kg	$0.291	$17.48	$146,846
Cellulase Complex	0	kg	$3.000	$0.00	$0
Natural Gasoline (stm 701)	0	kg	$0.155	$0.00	$0
Diesel/Gasoline	3	kg	$0.155	$0.46	$3,899
WWT Chemicals	0.0	kg	$2.237	$0.00	$0
CW Chemicals	0.0	kg	$1.428	$0.00	$0
BFW Chemicals	0.0	kg	$0.226	$0.00	$0
Boiler Fuel (stm 813)	2	Mbtu	$2.500	$6.22	$52,227
Total Raw Materials				$71	$593,315

Processing Material Costs

$per lb. calcs.

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Antifoam (Corn Oil)	79	kg	$0.441	$35	$291,248
Total Processing Materials				$35	$291,248

Operations and Maintenance Costs - DRY HANDLING (area 100)	each/day		wage	hr/day each	Total Cost /yr.
Supervisors	0.019	$	20.00	12	$1,650
Operators	0.075	$	16.00	12	$5,281
Laborers	0.301	$	16.00	12	$21,124
Maintenance	0.075	$	16.00	12	$5,281
Operations and Maintenance Costs - HYDROLYSIS/FERMENTATION (area 200, 300, 400, 500, 600)					
Supervisors	0.0	$	20.00	12	$1,650
Operators	0.1	$	16.00	8	$5,281
Laborers	0.0	$	16.00	8	$1,760
Technicians (Includes Lab.)	0.1	$	16.00	8	$5,281
Maintenance	0.1	$	16.00	8	$5,281
Operations and Maintenance Costs - Utilities (area 700, 800, 900)					
Supervisors	0.0	$	20.00	12	$825
Operators	0.1	$	16.00	8	$2,641
Laborers	0.0	$	16.00	8	$880
Technicians	0.0	$	16.00	8	$880
Maintenance	0.1	$	16.00	8	$1,760
Total Operations and maintenance labor costs					$59,577

Other Operations and Maintenance Costs			
Payroll Overhead	35% of operating labor	$	20,852
Maintenance Costs	2% of plant cost	$	145,294
Operating Supplies	0.25% of plant cost	$	18,162
Environmental	0.50% of plant cost	$	36,324
Local Taxes	1% of plant cost	$	72,647
Insurance	0.50% of plant cost	$	36,324
Overhead Costs	40% of labor, supervision,maint cost	$	23,831
Administrative Costs	1% of annual sales (less tax credits)	$	3,929
Distribution and Sales	0.5% of annual sales (less tax credits)	$	-
Total O&M Costs			$416,939

Operating Expenses:			
Utilities		1,757,995	2,349,741
Raw Materials		593,315	796,588
Processing Materials		291,248	267,948
Operation & Maintenance		416,939	680,780
Property Tax @	0.50% Book Value	45,656	79,234
Depreciation		608,749	1,056,454
Debt retierment		1,254,118	1,418,471
Total Operating Expense ($/yr)		**$4,968,020**	6,649,217

Savings With PureVision $ 1,681,197 / yr = 25.3%

based only on estimated enzyme production costs

ASSUMPTIONS

		NREL (.45)	3442
Fraction of pre-treated liquor required		5.024%	3.768%
Fraction of wastewater treated		4.38%	3.28%
Fraction of steam required		1.74%	1.31%
Fraction of ammonia required		15.77%	15.54%
Fraction of CSL required		22.03%	17.57%
ammonia storage tank estemated cost	$	100,000	

Comparison of On-site and Purchased Cellulase

Comparison of On-Site Cellulase Production via Pure Vision Technology and NREL Reference Model, to Purchase of Commercially Available Enzyme

CURRENT ASSUMPTION: ***BASED ON PUREVISION LABORATORY RESULTS OF COMPARISON***

	NREL* M FPU required/yr**	difference	Pure Vision M FPU required/yr	difference	Purchased Cellulase *** M FPU required/yr
	1,446,984	(50,708)	1,497,692	56,431	1,554,123
perating Projection:					
gal of fuel grade ethanol produced	$ 25,434,849	$ (311,275)	$ 25,746,124	$ 933,825	$ 26,679,948
Contract sale price per gallon	$ 1	$ -	$ 1	$ -	$ 1
Gross Annual Revenue	$ 27,978,334	$ (342,402)	$ 28,320,736	$ 1,027,207	$ 29,347,943
Small Ethanol Producer Tax Credit					
@ $ - per gallon	$ -		$ -		$ -
Total projected ethanol sales and credit	$ 27,978,334	$ (342,402)	$ 28,320,736	$ 1,027,207	$ 29,347,943
Gross Annual Co-Product Revenue	$ 328,822	$ -	$ 328,822	$ -	$ 328,822
Gross Sales and Credit	$ 28,307,156	$ (342,402)	$ 28,649,558	$ 1,027,207	$ 29,676,765
Operating Expenses:					
Utilities	$ 4,792,171	$ 567,400	$ 4,224,771	$ (1,803,557)	$ 2,421,214
Raw Materials	$ 12,843,241	$ 96,523	$ 12,746,718	$ 4,488,530,135	$ 4,501,276,853
Processing Materials	$ 267,948	$ 66,987	$ 200,961	$ (200,961)	$ -
Operation & Maintenance	$ 6,414,114	$ 70,428	$ 6,343,686	$ (505,618)	$ 5,838,069
Property Tax @ 0.50% Book Value	$ 486,736	$ 57,315	$ 429,421	$ (28,534)	$ 400,888
Depreciation	$ 6,038,644	$ 744,902	$ 5,293,743	$ (340,048)	$ 4,953,694
Total Operating Expense	$ 30,842,855	$ 1,603,554	$ 29,239,301	$ 4,485,651,417	$ 4,514,890,718
Net Operating Income	$ (2,535,699)	$ (1,945,956)	$ (589,742)	$(4,484,624,210)	$ (4,485,213,953)
				$ -	
Net Operating Cash Flow	$ 3,502,945	$ (1,201,055)	$ 4,704,000	$(4,484,964,258)	$ (4,480,260,258)

		NREL	Pure Vision	Purchased Cellulase
enzyme cost (cost of production calculated in "$per lb. calcs.") divided by lbs. per year flow rate from mass balance.	$/lb	$ 0.027	$ 0.020	$ 2.413
enzyme cost (cost of production calculated in "$per lb. calcs.") divided by million FPU per year required.	**$/MFPU**	**$ 4.60**	**$ 3.32**	**$ 2,753.93**

Annual Savings Using PureVision On-Site Enzyme Production	
OVER REFERENCE MODEL:	$ 1,201,055
OVER PURCHASED ENZYME:	$ 4,484,964,258

* 45% scale factor applied, SHCF

** MFPU = million FPU

*** Specialty Enzymes, Liquicell 2500, $2.00/lb, S.G. 1.100, 32 FPU/ml.

PLAINS YORK MODEL WITH PURCHASED CELLULASE FOR COMPARISON OF ON-SITE ENZYME PRODUCTION VS. PURCHASED

GAIN IN ETOH PRODUCTION POSSIBLE: 332 kg/hr

A

10/27/99

_ ENZYMATIC HYDROLYSIS - PRO FORMA

lying Assumptions & Input Variables

JRRENT SITUATION:

The Pro Forma models an Enzymatic Hydrolysis Ethanol plant using corn stover as the feed stock.

ETHANOL

The plant will convert corn stover to fuel grade ethanol utilizing enzymatic hydrolosis.

Corn stover feed rate of 71,977 kg/hr (str 101), produce estimated total output in
equivalent kilograms of fuel grade ETOH 9,483 kg/hr. = 79,659,865 kg / year (str 515)
gal./short ton= 76.8 3,176 gal/hr = 26,679,948 gal / year
gal./metric ton= 84.7

Increase to current York yearly production: 72%

The model assumes renewal of the ethanol excise tax credit of $.54 per gallon to the blender and **NOT** the small producer tax credit of $.10 per gallon through the year 2015 for a total ethanol value of
$1.10 per gallon or $0.37 per kg and **$ 29,347,943** **per year TOTAL Ethanol sales**

CARBON DIOXIDE

Currently, carbon dioxide from the High Plains York fermentations is sold to a CO_2 compression company.
Diverting the CO_2 (stm 550) from the stover plant into this stream for sale as opposed to the atmosphere provides
110,749 kg/hr = 930,294 ton / year with a value of $ 4.13 per metric ton

WITH THIS PROFORMA NO CO_2 IS SOLD. **CO_2 Value/year = $0**

LIGNIN

A Lignin co-product is produced and sold as combustion fuel material. A total amount of lignin in the stream (stm 601B) is
63,778 kg/hr = 535,734 metric ton / year is produced from the process.
The water in the lignin stream must be vaporized at a net BTU cost for the stream (stm 601B). Water vaporized is
43,969 kg/hr = 369,337 metric ton/year is vaporized at 1,100 BTU/lb loss = (107) MM BTU/hr
The remaining 19,809 kg/hr of stream 601B has 24,251 BTU/kg value = 480 MM BTU/hr
Total heating value from stream 601A is 374 MM BTU/hr

Gross Lignin Value/year = $7,848,926
Transport Cost = $7,848,926
Net Lignin Value = $0

METHANE

The digester produces 85% methane @ 353 kg/hr (stm 615) 44,332 BTU/kg CH4
Total heating value from Methane is 16 MM BTU/hr
methane is used in the DDG dryers and based on BTU value of $2.50 MM BTU

METHANE Value/year = $328,822

DIGESTER SLUDGE

The digester produces (stm 623) 0 kg/hr of sludge as fuel = 2,254 BTU/lb
based on 9,845 btu/lb biomass and 70% water in the sludge. = 4,969 BTU/kg
Total heating value from sludge is 0.00 MM BTU/hr

SLUDGE Value/year = $0

Sale of methane and lignin, based on BTU value is **$328,822** **per year**

Total projected facility sales would be **$29,676,765** per year

APITAL INVESTMENT ASSUMPTIONS

Total capital investment			
Civil Structural		(500,000)	
Area 100		6,146,434	
Area 200		14,955,166	
Area 300		4,028,307	
Area 307		3,714,334	
Area 400		651,440	
Area 500		7,515,486	
Area 600		9,824,251	
Area 700		234,910	
Area 800		3,684,612	
Area 900		2,236,491	
Fixed Capital		$52,491,432	
INDIRECTS Prorateable	3.5%	$1,837,200	
Process Development	2.0%	$1,049,829	
Field Expense	8.0%	$4,199,315	
Home Office Constr. Fee	12.0%	$6,298,972	
Contingency	10.0%	$5,249,143	
Start-up, Permits, Fees	3.0%	$1,574,743	
Working Capital per estimate		$375,592,910	1 mos Raw matls. + O&M
	Total Plant Cost	$448,293,544	
FEDERAL & STATE GRANTS	10%	($44,829,354)	
	Net Capital Investment	**$403,464,190**	

PERATING COST ASSUMPTIONS 8,400 hr/yr

Utilities (Rates based on 26,679,948 gal/yr produced)

	Amount/hr	*Units*	*$/unit*	*Cost /hr.*	**Total Cost /yr**
*Electricity	6,759	Kw-hr	$0.035	$237	$1,987,079
Well water	79,972	kg	$0.000	$0	$0
*Wastewater	39,119	kg	$0.00026	$10	$86,808
*Gypsum waste disposal	1,137	kg	$0.0364	$41	$347,327
		mTon	$1.103	$0	$0
Total Utilities				$288	$2,421,214

Quoted by High Plains

Model Input (purchased)

Raw Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Corn Stover DRY (stm 101 less water)	37,500	kg	$0.680	$25,499.90	$214,199,143
*Sulfuric Acid (stm 710)	860	kg	$0.100	$86.26	$724,592
*Calcium Hydroxide (Lime stm 227)	337	kg	$0.293	$98.70	$829,039
*Ammonia (stm 717)	387	kg	$0.162	$62.77	$527,281
Corn Steep Liquor (stm 735)	708	kg	$0.051	$36.10	$303,280
Nutrients (stm 415)	0	kg	$0.291	$0.00	$0
Purchased Cellulase	211,123	lbs	$2.000	$422,246.70	$3,546,872,248
transport cost	750	miles	$3.000	$2250 /load	$733,071,990
*Natural Gasoline (stm 701)	391	kg	$0.155	$60.36	$506,988
*Rolling Stock Gasoline	79	kg	$0.155	$12.32	$103,470
*WWT Chemicals	5	kg	$0.000	$0.00	$0
*CW Chemicals	17	kg	$0.000	$0.00	$0
*BFW Chemicals	73.8	kg	$0.226	$16.65	$139,833
*Boiler Fuel (stm 813)	190	Mbtu	$2.500	$476.07	$3,998,989
Total Raw Materials				$448,596	$4,501,276,853

**** Quoted by High Plains***

Processing Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
*Antifoam (Corn Oil)	0	kg	$0.304	$0	$0
Total Processing Materials				$0	$0

**** Quoted by High Plains***

Operations and Maintenance Costs - DRY HANDLING (area 100)	each/day		wage	hr/day, each	Total Cost /yr.
*Supervisors	0.5	$	20.00	12	$43,800
*Operators	2.0	$	16.00	12	$140,160
*Laborers	8.0	$	16.00	12	$560,640
*Maintenance	2.0	$	16.00	12	$140,160
Operations and Maintenance Costs - HYDROLYSIS/FERMENTATION (area 200, 300, 400, 500, 600)					
*Supervisors	1.0	$	20.00	12	$87,600
*Operators	8.0	$	16.00	8	$373,760
*Laborers	4.0	$	16.00	8	$186,880
*Technicians (Includes Lab.)	3.0	$	16.00	8	$140,160
*Maintenance	3.0	$	16.00	8	$140,160
Operations and Maintenance Costs - Utilities (area 700, 800, 900)					
*Supervisors	0.5	$	20.00	12	$21,900
*Operators	3.0	$	16.00	8	$70,080
*Laborers	1.0	$	16.00	8	$23,360
*Technicians	1.0	$	16.00	8	$23,360
*Maintenance	2.0	$	16.00	8	$46,720

**** Quoted by High Plains*** Standard HPY shifts are 12 hours.

Total Operations and maintenance labor costs $1,998,740

Other Operations and Maintenance Costs

Payroll Overhead	35% of operating labor	$	699,559
Maintenance Costs	2% of plant cost	$	1,049,829
Operating Supplies	0.25% of plant cost	$	131,229
Environmental	0.50% of plant cost	$	262,457
Local Taxes	1% of plant cost	$	524,914
Insurance	0.50% of plant cost	$	262,457
Overhead Costs	40% of labor, supervision,maint cost	$	799,496
Administrative Costs	1% of annual sales (less tax credits)	$	109,388
Distribution and Sales	0.5% of annual sales (less tax credits)	$	-
Total O&M Costs			$5,838,069

THER MODEL ASSUMPTIONS

/erage prevailing market price of fuel grade ETOH:		$0.37	per kg
ssumes renewal of the ethanol excise tax credit of $.54 per gallon		$ 1.10	per gallon
d the small producer tax credit of $.10 per gallon through the year 2007			
/alue of CO_2 produced		$ 4.13	per metric ton
Price for Electricity		$ 0.035	per KWhr
Gas price per million BTU		$ 2.500	per MM BTU
		68%	Dry matter
orn Stover feedstock cost- **dry basis/short ton**	$ 14.45	$0.016	per kg
		$15.93	per metric ton
ant on-stream factor		0.959	
ant operating hours per year		8,400	
epreciable Life of Capital Equipment		15	years
/erage annual commodity escalation rate:		3.0%	
/erage annual cost escalation rate:		3.0%	

*** Quoted by High Plains**

There are no land acquisiton costs included.
There are no off site costs included (e.g. public road improvements, extensions of power, water, telephone services)
There is a source of qualified construction personnel within daily driving distance of the site
There exist adequate roads and rail roads to allow equipment delivery.
The costs for air and water permits are not included.
Soils are adequate for conventional foundation designs.

CALCULATIONS FOR REQUIRED AMOUNT OF PURCHASED CELLULASE LIQUICELL 2500

ASED ON PUREVISION LABORATORY RESULTS OF COMPARISON

High Grade Waste Paper Substrate			
Soluble Charbohydrate % degraded in 18 hrs.			
Liquicell 2500	13%	87,059,020	ml/hr required for stover
PureVision Cellulase	82%	13,057,632	ml/hr required for stover
effectiveness multiple	6.43		

125	FPU/g protein	Liquicell 2500
731,295,772	liters/yr	Specialty
1.1000	S.G.	Enzymes
804,425,349	kg/yr	Inc.
193,062,084	gal/yr	
1,773,436,124	#/yr	
325,810	loads/yr	

1

cellulase storage tank

22,984	gal/hr
750,000	gal/vessel
33	vessel res. time (hr)

cellulase transfer pump

383 gpm

ASED ON PRODUCT SPECIFICATIONS PROVIDED BY SPECIALTY ENZYMES INC.

32	FPU/ml	Liquicell 2500
48,566,337	liters/yr	Specialty
1.1000	S.G.	Enzymes
53,422,971	kg/yr	Inc.
12,821,513	gal/yr	
117,776,282	#/yr	
21,637	loads/yr	

0

cellulase storage tank

14,021	gal/hr
750,000	gal/vessel
53	vessel res. time (hr)

cellulase transfer pump

234 gpm

ransport Calculations

10,000	lbs/axel		9.19	cellulase lb/gal
5	axels/truck		5,443	gal/truck
50,000	lbs/truck	$	**0.413**	**transport cost/lb**

stimated Equipment Costs for Reference Model Scaled Down 45% with Ranges to Equipa automatically update these cells with the exception of those noted in red.

Equip No.	No. Req'd	No. Spare	Equip Name	Scaling Stream	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Install Factor	Installed Cost	Scaled Uninstalled Cost in 1999$	Description	3442 WORK	
C-101	1	0	Bale conveyor	AREA0100	154	170	1.11	$15,000	1999	$15,000	0.6	$15,927	1.5	$24,551	$ 15,927	wire mesh conveyor 60" wide 20' long	WC101	11.93
C-102	1	0	Radial Stacker Conveyor	AREA0100	154	170	1.11	$159,830	1999	$159,830	0.6	$169,708	1.5	$261,604	$ 169,708	16 degree, 36" x 200' radial stacker, 750 ton/hr, 75 HP	WC102	44.74
C-103	1	0	Breaker Infeed Belt	AREA0100	154	170	1.11	$49,500	1999	$49,500	0.6	$52,559	1.5	$81,020	$ 52,559	84" x 35' rubber belt cleated infeed conveyor, 10 HP, TEFC drive motor with guard	WC103	5.97
C-104	1	0	1st Shredder Conveyor	AREA0100	154	170	1.11	$25,650	1999	$25,650	0.6	$27,235	1.5	$41,983	$ 27,235	60" wide x 25' long, 10 HP, TEFC drive with guard	WC104	5.97
C-105	1	0	1st Infeed Belt	AREA0100	154	170	1.11	$38,500	1999	$38,500	0.6	$40,879	1.5	$63,015	$ 40,879	60" wide x 30' long, 10 HP, TEFC drive with guard	WC105	11.93
C-106	1	0	2nd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.5	$48,285	$ 31,323	48" wide x 20' long, 7.5 HP, TEFC drive with guard	WC106	4.47
C-107	1	0	2nd Infeed Belt	AREA0100	154	170	1.11	$27,500	1999	$27,500	0.6	$29,200	1.5	$45,011	$ 29,200	48" wide x 30' long, 5 HP, TEFC drive with guard	WC107	2.98
C-108	1	0	3rd Shredder Conveyor	AREA0100	154	170	1.11	$29,500	1999	$29,500	0.6	$31,323	1.5	$48,285	$ 31,323	48" wide x 20' long, 10 HP, TEFC drive with guard	WC108	5.97
C-109	1	0	Feed Screw Conveyor	AREA0100	225,140	562,850	2.50	$31,700	1997	$31,700	0.6	$54,932	1.5	$86,351	$ 56,018	14" dia. 250' long	WC109	53.75
M-101	2	0	Truck Scale	AREA0100	96	72	0.75	$10,000	1999	$20,000	0.6	$16,829	1.5	$25,244	$ 16,829	96 deliveries /scale/12hr		
M-102	1	0	Receiving Pad	AREA0100	250,000	250,000	1.00	$2,083,500	1999	$2,083,500	0.6	$2,083,500	1.0	$2,083,500	$ 2,083,500	250,000 ft2 concrete pad, 9" thick with drainage		
M-103	6	1	Front End Loader	AREA0100	159,948	159,948	1.00	$156,000	1998	$1,092,000	0.6	$1,092,000	1.2	$ 1,326,016	$ 1,105,013	run on gasoline		
M-104	3	0	Bale Breaker	AREA0100	154	170	1.11	$250,000	1999	$750,000	0.6	$796,352	1.2	$955,622	$ 796,352	30 HP each	WM104	53.69
M-105	1	0	Primary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.2	$135,444	$ 112,870	250 HP, 1200 rpm, hammermill	WM105	149.14
M-106	1	0	Secondary Stover Shredder	AREA0100	154	170	1.11	$106,300	1999	$106,300	0.6	$112,870	1.5	$169,304	$ 112,870	250 HP, 1200 rpm, hammermill	WM106	149.14
M-107	1	0	Shred Bunker	AREA0100	600,000	600,000	1.00	$700,000	1999	$700,000	0.6	$700,000	1.0	$700,000	$ 700,000	200x100x30ft bunker with three walls, 3 days shred storage		
M-108	1	0	Storm Runoff Pond	AREA0100	1,747,767	1,747,767	1.00	$51,198	1998	$51,198	0.6	$51,198	1.0	$51,198	$ 51,808	200 x 150 x 8 ft, 240,000ft3		
										weighted averages:	0.6		1.1					499.68
A100									Subtotal	$5,315,978		$5,418,705		$6,146,434	$5,433,414			
							2000tpd x .45 (current year cost with area weighted-average scale exponent applied)						1.3	$3,181,636	($2,964,798)	is installed cost savings		
											Cost Base Year =	$1,999						
A-201	1	0	In-line Sulfuric Acid Mixer	STRM0214	55,308	23,725	0.43	$1,900	1997	$1,900	0.48	$1,266	1.2	$1,585	$1,291	Static Mixer, 110 gpm total flow		
A-202	1	0	In-line NH3 Mixer	STRM0244	53,630	18,317	0.34	$1,500	1997	$1,500	0.48	$896	1.2	$1,122	$913	Static Mixer, 82 gpm total flow		
A-209	1	0	Overliming Tank Agitator	STRM0228	167,050	102,608	0.61	$19,800	1997	$19,800	0.51	$15,442	1.2	$19,345	$15,748	Top Mounted, 1800 rpm, 15 hp	WT209	8.39
A-224	1	0	Reacidification Tank Agitator	STRM0239	167,280	102,752	0.61	$65,200	1997	$65,200	0.51	$50,851	1.2	$63,702	$51,857	Top-Mounted, 1800 rpm, 54 hp	WT224	25.17
A-232	1	0	Reslurrying Tank Agitator	STRM0250	358,810	167,795	0.47	$36,000	1997	$36,000	0.51	$24,432	1.2	$30,606	$24,915	Top-Mounted, 1800 rpm, 25 hp	WT232	13.98
A-235	1	0	In-line Acidification Mixer	STRM0236	164,570	101,104	0.61	$2,600	1997	$2,600	0.48	$2,058	1.2	$2,578	$2,099	Static-Mixer, 440 gpm total flow		
C-201	1	0	Hydrolyzate Screw Conveyor	STRM0220	225,140	101,493	0.45	$59,400	1997	$59,400	0.78	$31,908	1.5	$50,158	$32,539	18" dia. 33' long, 3420 cfh max flow, 23 hp	WC201	13.72
C-202	1	0	Wash Solids Screw Conveyor	STRM0225	196,720	165,453	0.84	$23,700	1997	$23,700	1	$19,933	1.5	$31,334	$20,327	18" dia. 16' long, 3420 cfh max flow	WC202	16.70
C-225	1	0	Lime Solids Feeder			0		$3,900	1997	$3,900	1	$3,900	1.5	$6,131	$3,977	6" dia., 63 cfh, 3150 lb/hr max flow	WC225	0.15
H-200	1	0	Hydrolyzate Cooler	AREA0200	1,988	895	0.45	$45,000	1997	$45,000	0.51	$29,947	2.2	$66,543	$30,539	Fixed Tube Sheet, 900 sf, 20" dia. X 20' long		
H-201	1	1	Beer Column Feed Economizer	AREA0201	5,641	5,641	1.00	$139,350	1999	$278,700	0.68	$278,700	2.2	$607,278	$278,700	TEMA type AES shell and tube 5641 sf, 42" dia x 20' long		
M-202	1	0	Prehydrolysis Reactor	STRM0217	270,034	121,514	0.45	$12,461,841	1998	$12,461,841	0.78	$6,684,746	1.5	$10,146,612	$6,764,408	Vertical Screw, 10 min residence time	WM105	353.16
P-201	1	1	Sulfuric Acid Pump	STRM0710	1,647	414	0.25	$4,800	1997	$9,600	0.79	$3,228	2.8	$9,190	$3,291	2 gpm, 245 ft. head	WP201	0.40
P-209	1	1	Overlimed Hydrolyzate Pump	STRM0228	167,050	102,608	0.61	$10,700	1997	$21,400	0.79	$14,561	2.8	$41,458	$14,849	448 gpm, 150 ft. head	WP209	18.01
P-222	1	1	Filtered Hydrolyzate Pump	STRM0230	162,090	101,614	0.63	$10,800	1997	$21,600	0.79	$14,936	2.8	$42,526	$15,231	448 gpm, 150 ft head	WP222	17.83
P-223	1	0	Lime Unloading Blower	STRM0227	547	337	0.62	$47,600	1998	$47,600	0.5	$37,340	1.4	$52,898	$37,785	3341 cfm, 6 psi, 10,024 lb/hr	WP223	4.10
P-224	1	1	Hydrolysis Feed Pump	STRM0250	160,000	167,795	1.05	$64,934	1999	$129,868	0.6	$133,628	1.2	$160,354	$133,628	740 gpm, 240 ft head	WP224	119.31
P-225	1	1	ISEP Elution Pump	STRM0243	52,731	18,005	0.34	$7,900	1997	$15,800	0.79	$6,761	2.8	$19,249	$6,894	104 gpm, 150 ft head	WP225	3.92
P-226	1	1	ISEP Reload Pump	STRM0246	164,080	100,802	0.61	$8,700	1997	$17,400	0.79	$11,841	2.8	$33,714	$12,075	445 gpm, 150 ft head	WP226	17.92
P-227	1	1	ISEP Hydrolyzate Feed Pump	STRM0221	160,290	98,157	0.61	$10,700	1997	$21,400	0.79	$14,526	2.8	$41,359	$14,814	432 gpm, 150 ft head	WP227	16.81
P-239	1	1	Reacidified Liquor Pump	STRM0239	167,280	102,752	0.61	$10,800	1997	$21,600	0.79	$14,698	2.8	$41,847	$14,988	450 gpm, 100 ft head	WP239	12.09
S-202	3	0	Pre-IX Belt Filter Press	SOLD0220	57,000	57,000	1.00	$200,000	1998	$600,000	0.39	$600,000	1.4	$850,010	$607,150	Use 3 units for 45% of the flow as recommended by the vendor	WS202	19.69
S-221	1	0	ISEP	STRM0240	210,005	98,157	0.47	$2,058,000	1997	$2,058,000	0.33	$1,601,194	1.2	$1,959,422	$1,632,851	10 chambers (39" dia. X 84" high), 4" dia. Valve - Weak Base Resin	WS221	2.98
S-222	1	0	Hydroclone & Rotary Drum Filter	STRM0229	5,195	1,137	0.22	$165,000	1998	$165,000	0.39	$91,224	1.4	$129,235	$92,311	Hydrocyclone and Vacuum Filter for 453 gpm	WS222	11.93
S-227	1	0	LimeDust Vent Baghouse	STRM0227	548	337	0.61	$32,200	1997	$32,200	1	$19,778	1.5	$30,254	$20,169	3750 cfm, 625 sf, 6 cfm/sf		
T-201	1	0	Sulfuric Acid Storage	STRM0710	1,647	860	0.52	$5,760	1996	$5,760	0.71	$3,633	1.7	$6,283	$3,751	2000 gal., 24 hr. residence time, 90% wv, 5.5ft diam. X 11ft		
T-203	1	0	Blowdown Tank	STRM0217	270,300	121,514	0.45	$64,100	1997	$64,100	0.93	$30,475	1.7	$52,061	$31,078	7000 gal., 11' dia x 30' high, 10 min. res. time, 75% wv, 15 psig		
T-209	1	0	Overliming Tank	STRM0228	167,050	102,608	0.61	$71,000	1997	$71,000	0.71	$50,232	1.8	$90,186	$51,225	29850 gal., 16' dia. X 32' high, 1 hr. res. time, 90% wv, 15 psig		
T-220	1	0	Lime Storage Bin	STRM0227	548	548	1.00	$69,200	1997	$69,200	0.46	$69,200	1.8	$124,243	$70,568	4455 cf, 14' dia x 25' high, 1.5x rail car vol., atmospheric, 15 day storage max		
T-224	1	0	Reacidification Tank	STRM0239	102,752	102,752	1.00	$111,889	1999	$111,889	0.51	$111,889	1.8	$196,992	$111,889	120,000 gal., 28' dia x 28' high, 4 hr. res. time, 90% wv, atmospheric		
T-232	1	0	Slurrying Tank	STRM0250	358,810	167,795	0.47	$44,800	1997	$44,800	0.71	$26,117	1.8	$46,890	$26,633	11300 gal., 13' dia. X 25' high, 15 min. res. time, 90% wv		
0	0	0	0	0	0	0	0.00	$0	1999	$0	0	$0	-	$0	$0	0		
										weighted averages:	0.696961		1.5					676.27
A200									Subtotal	$16,527,758		$9,999,337		$14,955,166	$10,128,493			
							2000tpd x .45 (current year cost with area weighted-average scale exponent applied)						1.5	$15,025,380	$70,213	is installed cost savings		

A-300	8	0	Fermentor Agitators	GALLONS	962,651	750,000	0.78	$19,676	1996	$157,408	0.51	$138,592	1.2	$175,799	$143,110	Side Mounted, 2 per vessel, 60 hp each, 0.15 hp/1000 gal	WT300	201.34
A-301	1	0	Seed Hold Tank Agitator	STRM0304	41,777	17,529	0.42	$12,551	1996	$12,551	0.51	$8,060	1.2	$10,223	$8,322	Top Mounted, 1800 rpm, 10 hp, 0.1 hp/1000 gal	WT301	5.59
A-304	2	0	4th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$11,700	1997	$23,400	0.51	$15,026	1.2	$18,824	$15,323	Top Mounted, 1800 rpm, 3 hp, 0.3 hp/1000 gal	WT304	3.36
A-305	2	0	5th Seed Vessel Agitator	STRM0304	41,777	17,529	0.42	$10,340	1996	$20,680	0.51	$13,280	1.2	$16,845	$13,713	Top Mounted, 1800 rpm, 9 hp, 0.1 hp/1000 gal	WT305	10.07
A-306	1	0	Beer Well Agitator	STRM0502	381,700	173,737	0.46	$10,100	1997	$10,100	0.51	$6,761	1.2	$8,469	$6,394	Top Mounted, 1800 rpm, 2 hp, 0.3 hp/1000 gal	WT306	1.12
F-300	4	0	Fermentors	GALLONS	750,000	750,000	1.00	$326,203	1999	$1,304,812	0.71	$1,304,812	1.8	$2,297,260	$1,304,812	750,000 gal. each, 2 day residence total, 90% wv, API, atmospheric, 50' f x 51'		
F-301	2	0	1st Fermentation Seed Fermentor	None		0	0.45	$14,700	1997	$29,400	0.93	$13,991	2.8	$39,948	$14,267	9 gal, jacketed, agitated, 1' dia., 1.5' high, 15 psig		
F-302	2	0	2nd Fermentation Seed Fermentor	None		0	0.45	$32,600	1997	$65,200	0.93	$31,027	2.8	$88,592	$31,640	90 gal., jacketed, agitated, 2' 3" dia., 3' high, 2.5 psig		
F-303	2	0	3rd Fermentation Seed Fermentor	None		0	0.45	$81,100	1997	$162,200	0.93	$77,186	2.8	$220,394	$78,712	900 gal., jacketed, agitated, 5' dia, 6.5' high, 2.5 psig		
F-304	2	0	4th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$39,500	1997	$79,000	0.93	$35,225	1.7	$60,174	$35,921	9000 gal., 9' dia x 19' high, atmospheric		
F-305	2	0	5th Fermentation Seed Fermentor	STRM0304	41,777	17,529	0.42	$147,245	1998	$294,490	0.51	$189,107	1.8	$336,910	$191,360	90000 gal., API, atmospheric 25'f x 25'		
H-300	4	1	Fermentation Cooler	QHX300EA	67,820	25,053	0.37	$4,000	1997	$20,000	0.78	$9,198	2.2	$20,438	$9,380	4 exchangers at 221 sf, U=300 BTU/hr sf F LMTD = 22.9°F plate and frame		
H-301	1	0	Fermentation Seed Hydrolyzate Cooler	AREA0301	773	318	0.41	$15,539	1998	$15,539	0.78	$7,778	2.2	$17,151	$7,871	348 sf, 300 BTU/hr sf F		
H-302	1	0	Fermentation Pre-Cooler	AREA0302	3,765	828	0.22	$25,409	1998	$25,409	0.78	$7,797	2.2	$17,193	$7,890	828 sf total, plate and frame		
H-304	1	0	4TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$3,300	1997	$3,300	0.83	$1,580	1.2	$1,934	$1,611	12 sf, 1" sch 40 pipe, 105 BTU/hr sf F		
H-305	1	0	5TH Seed Fermentor Coils	QSDF0301	38,339	15,789	0.41	$18,800	1997	$18,800	0.98	$7,881	1.2	$9,644	$8,037	138 sf, 2" sch 40 pipe, 92 BTU/hr sf F		
P-300	4	1	Fermentation Recirc./Transfer Pump	QHX300EA	67,737	55,505	0.82	$8,000	1997	$40,000	0.79	$34,177	2.8	$97,307	$34,852	844 gpm @ 150 ft sized based on heating rate	WP300	104.49
P-301	1	1	Fermentation Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$22,194	1998	$44,388	0.7	$24,168	1.4	$34,238	$24,456	280 gpm @ 150 ft head	WP301	5.95
P-302	2	0	Seed Transfer Pump	STRM0304	41,777	17,529	0.42	$54,088	1998	$108,176	0.7	$58,898	1.4	$83,440	$59,600	504 gpm total, 252 gpm each, 100 ft head	WP302	7.14
P-306	1	1	Beer Transfer Pump	STRM0502	381,701	173,737	0.46	$17,300	1997	$34,600	0.79	$18,579	2.8	$52,899	$18,947	790 gpm each, 171 ft head	WP306	34.47
T-301	1	0	Fermentation Seed Hold Tank	STRM0304	41,777	17,529	0.42	$161,593	1998	$161,593	0.51	$103,767	1.8	$184,870	$105,003	105000 gal., API atmospheric		
T-306	1	0	Beer Well	STRM0502	129,000	183,467	1.42	$111,889	1999	$111,889	0.51	$133,906	1.8	$235,756	$133,906	192,518 gal., 32' dia x 32' high, 4 hr. res. time, 95% wv, atmospheric		
										weighted averages:	0.6843466		1.8					
A300									Subtotal	$2,742,935		$2,240,795		$4,028,307	$2,255,629			373.53
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.3	$8,218,509	$4,190,202	is installed cost savings		
A-307	8	0	Enzymatic Hydrolysis Tank Agitators	STRM0302B	157,136	157,136	1.00	$19,676	1996	$157,408	0.51	$157,408	1.2	$199,666	$162,539	two side mounted 75 hp agitators / tank, 0.4hp/1000 gal.	WT307	251.67
H-307	12	0	Enzymatic Hydrolysis Tank Heater	STRM0302B	157,136	157,136	1.00	$15,000	1999	$180,000	0.78	$180,000	2.2	$392,214	$180,000	65 ft2 double pipe		
H-308	1	0	Pre-hydrolyzate cooler	STRM0302	145,536	145,536	1.00	$25,000	1999	$25,000	0.78	$25,000	2.2	$54,474	$25,000	481 ft2, parallel double pipe		
P-308	8	1	Hydrolyzer Bottoms Pump	STRM0302B	157,136	157,136	1.00	$121,690	1999	$1,095,210	0.6	$1,095,210	1.2	$1,314,252	$1,095,210	3000 GPM each Disc flow pumps, 245ft head	WP308	1,744.94
T-307	4	0	Enzymatic Hydrolysis Tank	STRM0302B	750,000	375,000	0.50	$326,203	1999	$1,304,812	0.6	$860,855	2.0	$1,753,728	$860,855	375,000 gallons, 24 hour residence time, 2 side mounted agitators cone bottom, concrete base, bottom outlet through the concrete, 30o cone bottom		
0	0	0	0	0	0	0	0.00	$0	1999	$0	0	$0	-	$0	$0	0		
										weighted averages:	0.6082295		1.6					1,996.61
Area 307									Subtotal	$2,762,430		$2,318,473		$3,714,334	$2,323,604			
												2000tpd x .45 (cu	-	$0	($3,714,334)	is installed cost savings		
																assumed to be adequate equipment for distribution and storage of purchased enzyme		
P-420	1	1	Cellulase Transfer Pump (assumed same as reference model recycle water pump)	STRM0602	179,446	84,120	0.47	$10,600	1997	$21,200	0.79	$11,652	2.8	$33,175	$11,882	370 gpm, 150ft head	WP630	14.75
A-401	2	0	Cellulase Storage Tank Agitators (assumed same as study model fermentor agitators)	GALLONS	962,651	750,000	0.78	$19,676	1996	$39,352	0.51	$34,648	1.2	$43,950	$35,777	Side Mounted, 2 per vessel, 60 hp each, 0.15 hp/1000 gal	WT401	67.11
T-706	1	0	Cellulase Storage Tank (assumed same as study model production fermenter)	GALLONS	750,000	750,000	1.00	$326,203	1999	$326,203	0.71	$326,203	1.8	$574,315	$326,203	750,000 gal., 34 hr supply by purchase projection method "A" or 42 hr supply by purchase projection method "B", API, atmospheric, 50' ϕ x 51'		
												area install factor	1.7					81.87
A400									Subtotal	$386,755		$372,503		$651,440	$373,863			
												2000tpd x .45	1.3	$7,057,277	$6,405,837			

D-501	1	0	Beer Column	DIAMD501	4	2.29	0.56	$636,976	1996	$636,976	0.78	$402,792	2.1	$873,434	$415,921	7'6" DIA, 32 ACTUAL TRAYS, NUTTER V-GRID TRAYS		
D-502	1	0	Rectification Column	S510S521	56,477	26,744	0.47	$525,800	1996	$525,800	0.78	$293,491	2.1	$636,421	$303,058	8' dia.(rect)., 4' dia. (strip) x 18" T.S., 60 act. Trays, 60% eff., Nutter V-Grid trays		
E-501	1	0	1st Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,676	1996	$435,676	0.68	$435,676	2.1	$944,742	$449,877	22278 sf each., 135 BTU/hr sf F		
E-502	1	0	2nd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.1	$944,685	$449,850	22278 sf., 170 BTU/hr sf F		
E-503	1	0	3rd Effect Evaporation	AREA0502	22,278	22,278	1.00	$435,650	1996	$435,650	0.68	$435,650	2.1	$944,685	$449,850	22278 sf each., 170 BTU/hr sf F		
H-501	1	0	Beer Column Reboiler	QRFD0501	-7,863,670	-3,723,722	0.47	$158,374	1996	$158,374	0.68	$95,263	2.2	$214,340	$98,368	Fixed TS, 6602 sf, 31" dia., 20' long, 178 BTU/hr sf F		
H-502	1	0	Rectification Column Reboiler	QRFD0502	-987,427	-467,581	0.47	$29,600	1997	$29,600	0.68	$17,805	2.2	$39,563	$18,157	Thermosyphon, 512 sf, 15" dia., 20' long, 130 BTU/hr sf F		
H-504	1	0	Beer Column Condenser	QCND0501	277,820	131,557	0.47	29,544	1996	$29,544	0.68	$17,771	2.2	$39,984	$18,350	Floating Head, 418 sf, 15" dia., 22' long, 92 BTU/hr sf F		
H-505	1	0	Rectification Column Condenser	QCND0502	4,905,410	2,322,883	0.47	86,174	1996	$86,174	0.68	$51,834	2.2	$116,626	$53,524	Fixed TS, 1969 sf, 29" dia, 20' long, 157 BTU/hr sf F		
H-512	1	1	Beer Column Feed Interchange	AREA0512	909	430	0.47	$19,040	1996	$38,080	0.68	$22,905	2.2	$51,537	$23,652	431 sf, 200 BTU/hr sf F		
H-517	1	1	Evaporator Condenser	QHET0517	6,764,222	3,203,095	0.47	$121,576	1996	$243,152	0.68	$146,257	2.2	$329,077	$151,024	Fixed TS, 3906 sf, 29" dia., 20' long, 220 BTU/hr sf F		
M-503	1	0	Molecular Sieve (9 pieces)	STRM0515	20,491	9,703	0.47	$2,700,000	1998	$2,700,000	0.7	$1,599,964	1.0	$1,619,030	$1,619,030	Superheater, twin mole sieve columns, product cooler, condenser, pumps, vacuum source.	WM503	55.00
P-501	1	1	Beer Column Bottoms Pump	P501FLOW	5,053	2,200	0.44	$42,300	1997	$84,600	0.79	$43,861	2.8	$124,881	$44,728	2200 gpm, 150 ft head	WP501	84.65
P-503	1	1	Beer Column Reflux Pump	QCND0501	277,820	131,557	0.47	$1,357	1998	$2,714	0.79	$1,504	2.8	$4,248	$1,522	6 gpm, 140 ft head	WP503	0.22
P-504	1	1	Rectification Column Bottoms Pump	STRM0516	31,507	15,530	0.49	$4,916	1998	$9,832	0.79	$5,622	2.8	$15,884	$5,689	76 gpm, 158 ft head	WP504	2.80
P-505	1	1	Rectification Column Reflux Pump	QCND0502	4,906,301	2,323,304	0.47	$4,782	1998	$9,564	0.79	$5,299	2.8	$14,970	$5,362	207 gpm, 110 ft head	WP505	5.14
P-511	2	1	1st Effect Pump	STRM0525	278,645	133,617	0.48	$19,700	1997	$59,100	0.79	$33,069	2.8	$94,155	$33,723	1137 gpm each, 110 ft head	WP511	67.89
P-512	1	1	2nd Effect Pump	STRM0528	91,111	45,390	0.50	$13,900	1997	$27,800	0.79	$16,032	2.8	$45,646	$16,349	599 gpm, 110 ft head	WP512	17.37
P-513	2	1	3rd Effect Pump	STRM0531	48,001	23,814	0.50	$8,000	1997	$24,000	0.79	$13,795	2.8	$39,276	$14,068	196 gpm each, 110 ft head	WP513	12.54
P-514	1	1	Evaporator Condensate Pump	STRM534A	140,220	69,285	0.49	$12,300	1997	$24,600	0.79	$14,095	2.8	$40,131	$14,374	293 gpm, 125 ft head	WP514	9.20
P-515	1	1	Scrubber Bottoms Pump	STRM0551	15,377	7,427	0.48	$2,793	1998	$5,586	0.79	$3,143	2.8	$8,881	$3,181	31 gpm, 104 ft head	WP515	0.84
P-517	1	1	Kill Tank Bottoms Pump	STRM0518	5,053	660	0.13	$42,300	1997	$84,600	0.79	$16,944	2.8	$48,242	$17,279	660gpm, 72 ft head	WP517	12.19
T-503	1	0	Beer Column Relfux Drum	QCND0501	277,820	131,557	0.47	$11,900	1997	$11,900	0.93	$5,938	1.7	$10,144	$6,055	164 gal, 15 min res. Time, 50% wv, 2'6" dia., 5' long, 25 psig		
T-505	1	0	Rectification Column Reflux Drum	QCND0502	4,906,301	2,323,304	0.47	$45,600	1997	$45,600	0.72	$26,621	1.7	$45,476	$27,147	6225 gal, 15 min res time, 50% wv, 7' dia, 22' long, 25 psig		
T-512	1	0	Vent Scrubber	STRM0523	18,523	9,788	0.53	$99,000	1998	$99,000	0.78	$60,197	1.7	$102,043	$60,915	5' dia x 25' high, 4 stages, plastic Jaeger Tri-Packing		
T-513	1	0	Kill Tank	STRM0518	149,897	149,897	1.00	$99,920	1999	$99,920	0.78	$99,920	1.7	$167,384	$99,920	18 psig, 30 min. res. time		
										weighted averages:	0.7164992		1.7					267.85
A500									Subtotal	$6,343,492		$4,301,097		$7,515,486	$4,400,972			
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.7	$6,765,614	($749,372)	is installed cost savings		

C-601	1	0	Lignin conveyor	STRM0601B	225,140	225,140	1.00	$31,700	1997	$31,700	0.60	$31,700	1.5	$49,832	$32,327	14" dia. 100' long	WC109	21.50
M-613	1	0	Syrup Sprayer	STRM0531	22,372	22,372	1.00	$1,000	1999	$1,000	0.30	$1,000	1.2	$1,200	$1,000	100 GPM syrup sprayer		
M-614	1	0	Lignin Loadout	STRM0601A	63,778	0	0.00	$41,200	1999	$41,200	0.30	$0	1.0	$0	$0	245 GPM @ 20.6% insoluble solids		
M-615	1	0	Equalization Basin	STRM0830	98,267	102,204	1.04	$350,000	1999	$350,000	0.79	$361,031	1.0	$361,031	$361,031	no less than 500,000 gal., above-ground bolted tank with cover, including foundations, pumps and controls	WM615	1,077.21
M-616	1	0	Anaerobic Digestion System	STRM0830	98,267	102,204	1.04	$3,200,000	1999	$3,200,000	0.79	$3,300,852	1.0	$3,300,852	$3,300,852	500,000 gal., includes site work, foundations, reactors and ancillary equipment		
M-617	1	0	Aerobic Digestion System	STRM0830	98,267	102,204	1.04	$4,300,000	1999	$4,300,000	0.79	$4,435,520	1.0	$4,435,520	$4,435,520	four-350,000 gal. Sequencing Batch Reactors, 48,000 lbs/day of O2 transfer capability, de-nitrification facilities, aeration and mixing requires approximately 1,400 horsepower		
M-618	1	0	Pressure Sand Filters	STRM0830	98,267	102,204	1.04	$280,000	1999	$280,000	0.79	$288,825	1.0	$288,825	$288,825	400 ft2 of filtration surface area, includes the engineering and legal cost to acquire an NPDES permit		
P-630	1	1	Recycle Water Pump	STRM0602	179,446	84,120	0.47	$10,600	1997	$21,200	0.79	$11,652	2.8	$33,175	$11,882	370 gpm, 150ft head	WP630	14.75
S-601	2	0	Beer Column Bottoms Centrifuge	CENTFLOW	404	300	0.74	$659,550	1998	$1,319,100	0.60	$1,103,371	1.2	$1,339,824	$1,116,520	requires 540gpm duty, 2 @ 300 gpm and 410 hp each	WS601	489.18
T-630	1	0	Recycled Water Tank	STRM0602	179,446	84,120	0.47	$14,515	1998	$14,515	0.75	$8,254	1.7	$13,992	$8,353	7410 gal, 20 min. res., 2.5 psig, 9.5ft diam. x 14.25ft		
										weighted averages:	0.7609184		1.0					1,602.64
A600									Subtotal	$9,558,715		$9,542,206		$9,824,251	$9,556,310			
										2000tpd x .45 (current year cost with area weighted-average scale exponent applied)			1.3	$5,167,342	($4,656,910)	is installed cost savings		

P-703	1	1	Sulfuric Acid Pump	STRM0710	1,647	1,912	1.16	$8,000	1997	$16,000	0.79	$18,001	2.8	$51,253	$18,357	215 gpm, 150ft head	WP703	0.09
P-720	1	1	CSL Pump	STRM0735	2,039	859	0.42	$8,800	1997	$17,600	0.79	$8,889	2.8	$25,308	$9,065	182 gpm, 150ft head	WP720	0.15
T-703	1	0	Sulfuric Acid Storage Tank	STRM0710	1,647	1,912	1.16	$42,500	1997	$42,500	0.51	$45,860	1.8	$82,338	$46,767	20,000 gal, 240 hr supply, 90% wv, 12ft diam. x 24 ft, atmospheric		
T-720	1	0	CSL Storage Tank	STRM0735	2,039	859	0.42	$88,100	1997	$88,100	0.79	$44,495	1.7	$76,011	$45,375	30160 gal, 90% wv, 120 supply, 14.3ft diam. X 25 ft		
												area install factor	2.0					0.24
										$164,200				**$234,910**	**$119,563**			
								Subtotal				2000tpd x .45		$819,339	$584,429			
M-803	1	0	Boiler with Superheater	M0815 + 216	200,000	200,000	1	1,590,000	1999	$1,590,000	0.7	$1,590,000	1.3	$2,067,000	$1,590,000	200,000 #/hr running @ 171,488 #/hr; with 40,000 #/hr 160o superheat; 132,000#/hr 390o sat. @ 205 psig	WM803	75.60
M-820	1	0	Hot process water softener system	STRM0811B	229,386	45,003	0.20	$1,383,300	1999	$1,383,300	0.6	$520,623	1.2	$624,748	$520,623	200 gpm		
M-830	1	0	Hydrazine Addition Pkg.	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM830	10.00
M-832	1	0	Ammonia Addition Pkg	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM832	10.00
M-834	1	0	Phosphate Addition Pkg	STRM813A	229,386	80,536	0.35	$19,000	1994	$19,000	0.6	$10,139	1.0	$10,857	$10,857	75 gal tank, agitator, 2 metering pumps	WM834	10.00
P-804	2	1	Condensate Pump	STRM811A	249,633	38,798	0.16	$7,100	1997	$21,300	0.79	$4,894	4.6	$22,958	$4,991	130 gpm, 150' head	WP804	9.21
P-824	2	1	Deaerator Feed Pump	STRM811A	196,000	38,798	0.20	$9,500	1997	$28,500	0.79	$7,927	8.3	$67,097	$8,084	180 gpm, 115' head	WP824	4.89
P-826	4	1	BFW Pump	STRM0813	207,310	80,536	0	$52,501	1998	$262,505	0.79	$124,377	1.4	$176,203	$125,859	310 gpm, 2740' head	WP826	400.99
P-828	1	1	Blowdown Pump	STRM0821	6,600	2,699	0	$5,100	1997	$10,200	0.79	$5,032	6.4	$32,842	$5,132	12 gpm, 150' head	WP828	0.42
P-830	1	1	Hydrazine Transfer Pump	STRM813A	229,386	80,536	0	5,500	1997	$11,000	0.79	$4,811	6.4	$31,402	$4,907	3 gpm, 75' head	WP830	0.05
T-804	1	0	Condensate Collection Tank	STRM811A	229,386	38,798	0	7,100	1997	$7,100	0.71	$2,011	3.3	$6,766	$2,050	200 gal, 1.5 min. res. time		
T-824	1	0	Condensate Surge Drum	STRM811A	150,000	38,798	0.26	$49,600	1997	$49,600	0.72	$18,734	5.0	$95,523	$19,105	2100 gal., 6' diam. X 10', 15 psig, res. time 11 min.		
T-826	1	0	Deaerator	STRM0813	267,000	80,536	0.30	$165,000	1998	$165,000	0.72	$69,616	6.5	$457,896	$70,446	3030 gal., 15 psig, 10 min. res.		
T-828	1	0	Blowdown Flash Drum	STRM0821	6,550	2,699	0.41	$9,200	1997	$9,200	0.72	$4,859	7.3	$36,168	$4,955	210 gal., 2.5' diam. X 6', 50 psig 17 min. res.		
T-830	1	0	Hydrazine Drum	STRM813A	229,386	80,536	0.35	$12,400	1997	$12,400	0.93	$4,685	7.0	$33,440	$4,777	138 gal, 3.75' x 1.25' diam., 10 psig		
										weighted averages:	0.6704429		1.5					521.16
A800								**Subtotal**		**$3,607,105**		**$2,387,986**		**$3,684,612**	**$2,393,497**			
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.1	$23,046,972	$19,362,360	is installed cost savings		
M-902	1	0	Cooling Tower System	QCWCAPIT	41,100,000	12,955,985	0.32	$1,659,000	1998	$1,659,000	0.78	$674,181	1.2	$818,659	$682,216	40,000 gpm, 185.4MM BTU/hr	WM902	298.85
M-904	1	0	Plant Air Compressor	STRM0101	159,950	159,950	1.00	$60,100	1997	$60,100	0.34	$60,100	1.3	$79,675	$61,288	450 cfm, 125 psig outlet	WM904	186.40
M-908	1	0	Chilled Water Package	QCHLWCAP	5,040,000	2,268,000	0.45	$380,000	1997	$380,000	0.8	$200,610	1.2	$245,492	$204,577	1000 ton, 600kW	WM908	600.00
M-910	1	0	CIP System	STRM0914	63	28	0.45	$95,000	1995	$95,000	0.6	$58,837	1.2	$73,021	$60,851	designed by Delta-T, (est 0.2 kW)	WM910	0.20
P-902	1	1	Cooling Water Pumps	STRM0940	18,290,000	5,553,791	0.30	$332,300	1997	$664,600	0.79	$259,201	2.8	$737,993	$264,326	12300 gpm, 70ft head		
P-912	1	1	Make-up Water Pump	STRM0904	244,160	82,445	0.34	$10,800	1997	$21,600	0.79	$9,161	2.8	$26,084	$9,343	370 gpm, 75ft head	WP912	7.32
P-914	1	1	Process Water Circulating Pump	STRM0905	352,710	111,503	0.32	$11,100	1997	$22,200	0.79	$8,938	2.8	$25,449	$9,115	745 gpm, 75ft head	WP914	14.78
S-904	1	1	Instrument Air Dryer	STRM0101	159,950	71,977	0	$15,498	1999	$30,996	0.6	$19,197	1.3	$24,956	$19,197	134 scfm air dryer, -40F Dewpoint	WS601	4.91
T-904	1	0	Plant Air Receiver	STRM0101	159,950	53,316	0	$13,000	1997	$13,000	0.72	$5,894	1.7	$10,069	$6,011	300 gal., 200 psig		
T-914	1	0	Process Water Tank	STRM0905	352,710	111,503	0	195,500	1997	$195,500	0.51	$108,663	1.8	$195,095	$110,811	234360 gal, 8hr res. time		
																		1,112.46
Area 900								**Subtotal**		**$3,141,996**		**$1,404,783**		**$2,236,491**	**$1,427,733**		**Total kW**	**6,759**
								2000tpd x .45 (current year cost with area weighted-average scale exponent applied)					1.3	$5,278,320	$3,041,829	is installed cost savings		

3442 PLANT TOTAL:	**$50,551,364**	**$37,985,886**	**$52,991,432**
45% NREL TOTAL:			**$74,560,389**
SAVINGS:			**$21,568,956**
			28.93%

Comparison of On-Site Cellulase Production via Pure Vision Technology and NREL Reference Model, to Purchase of Commercially Available Enzyme

CURRENT ASSUMPTION: ***BASED ON PRODUCT SPECIFICATIONS PROVIDED BY SPECIALTY ENZYMES INC.***

		NREL* M FPU required/yr**	difference	Pure Vision M FPU required/yr	difference	Purchased Cellulase *** M FPU required/yr
		1,446,984	(50,708)	1,497,692	56,431	1,554,123
perating Projection:						
gal of fuel grade ethanol produced		$ 25,434,849	$ (311,275)	$ 25,746,124	$ 933,825	$ 26,679,948
Contract sale price per gallon		$ 1	$ -	$ 1	$ -	$ 1
Gross Annual Revenue		$ 27,978,334	$ (342,402)	$ 28,320,736	$ 1,027,207	$ 29,347,943
Small Ethanol Producer Tax Credit						
@ $ - per gallon		$ -		$ -		$ -
Total projected ethanol sales and credit		$ 27,978,334	$ (342,402)	$ 28,320,736	$ 1,027,207	$ 29,347,943
Gross Annual Co-Product Revenue		$ 328,822	$ -	$ 328,822	$ -	$ 328,822
Gross Sales and Credit		$ 28,307,156	$ (342,402)	$ 28,649,558	$ 1,027,207	$ 29,676,765
Operating Expenses:						
Utilities		$ 4,792,171	$ 567,400	$ 4,224,771	$ (1,803,557)	$ 2,421,214
Raw Materials		$ 12,843,241	$ 96,523	$ 12,746,718	$ 492,822,759	$ 505,569,478
Processing Materials		$ 267,948	$ 66,987	$ 200,961	$ (200,961)	$ -
Operation & Maintenance		$ 6,414,114	$ 70,428	$ 6,343,686	$ (505,618)	$ 5,838,069
Property Tax @ 0.50% Book Value		$ 486,736	$ 57,315	$ 429,421	$ (28,534)	$ 400,888
Depreciation		$ 6,038,644	$ 744,902	$ 5,293,743	$ (340,048)	$ 4,953,694
Total Operating Expense		$ 30,842,855	$ 1,603,554	$ 29,239,301	$ 489,944,041	$ 519,183,342
Net Operating Income		$ (2,535,699)	$ (1,945,956)	$ (589,742)	$ (488,916,834)	$ (489,506,577)
					$ -	
Net Operating Cash Flow		$ 3,502,945	$ (1,201,055)	$ 4,704,000	$ (489,256,883)	$ (484,552,883)
enzyme cost (cost of production calculated in "$per lb. calcs.") divided by lbs. per year flow rate from mass balance.	**$/lb**	$ 0.027		$ 0.020		$ 2.413
enzyme cost (cost of production calculated in "$per lb. calcs.") divided by million FPU per year required.	**$/MFPU**	**$ 4.60**		**$ 3.32**		**$ 182.89**

Annual Savings Using PureVision On-Site Enzyme Production	
OVER REFERENCE MODEL:	**$ 1,201,055**
OVER PURCHASED ENZYME:	**$ 489,256,883**

* 45% scale factor applied, SHCF

* * MFPU = million FPU

* * * Specialty Enzymes, Liquicell 2500, $2.00/lb, S.G. 1.100, 32 FPU/ml.

PLAINS YORK MODEL WITH PURCHASED CELLULASE FOR COMPARISON OF ON-SITE ENZYME PRODUCTION VS. PURCHASED
GAIN IN ETOH PRODUCTION POSSIBLE: 332 kg/hr

A
10/27/99

ENZYMATIC HYDROLYSIS - PRO FORMA
lying Assumptions & Input Variables

JRRENT SITUATION:

The Pro Forma models an Enzymatic Hydrolysis Ethanol plant using corn stover as the feed stock.

ETHANOL

The plant will convert corn stover to fuel grade ethanol utilizing enzymatic hydrolosis.
Corn stover feed rate of 71,977 kg/hr (str 101), produce estimated total output in
equivalent kilograms of fuel grade ETOH 9,483 kg/hr. = 79,659,865 kg / year (str 515)
gal./short ton= 76.8 3,176 gal/hr = 26,679,948 gal / year
gal./metric ton= 84.7

Increase to current York yearly production: 72%
The model assumes renewal of the ethanol excise tax credit of $.54 per gallon to the blender
and **NOT** the small producer tax credit of $.10 per gallon through the year 2015 for a total ethanol value of
$1.10 per gallon or $0.37 per kg and **$ 29,347,943 per year TOTAL Ethanol sales**

CARBON DIOXIDE

Currently, carbon dioxide from the High Plains York fermentations is sold to a CO_2 compression company.
Diverting the CO_2 (stm 550) from the stover plant into this stream for sale as opposed to the atmosphere provides
110,749 kg/hr = 930,294 ton / year with a value of $ 4.13 per metric ton
WITH THIS PROFORMA NO CO_2 IS SOLD. **CO_2 Value/year = $0**

LIGNIN

A Lignin co-product is produced and sold as combustion fuel material. A total amount of lignin in the stream (stm 601B) is
63,778 kg/hr = 535,734 metric ton / year is produced from the process.
The water in the lignin stream must be vaporized at a net BTU cost for the stream (stm 601B). Water vaporized is
43,969 kg/hr = 369,337 metric ton/year is vaporized at 1,100 BTU/lb loss = (107) MM BTU/hr
The remaining 19,809 kg/hr of stream 601B has 24,251 BTU/kg value = 480 MM BTU/hr
Total heating value from stream 601A is 374 MM BTU/hr
Gross Lignin Value/year = $7,848,926
Transport Cost = $7,848,926
Net Lignin Value = $0

METHANE

The digester produces 85% methane @ 353 kg/hr (stm 615) 44,332 BTU/kg CH4
Total heating value from Methane is 16 MM BTU/hr
methane is used in the DDG dryers and based on BTU value of $2.50 MM BTU
METHANE Value/year = $328,822

DIGESTER SLUDGE

The digester produces (stm 623) 0 kg/hr of sludge as fuel = 2,254 BTU/lb
based on 9,845 btu/lb biomass and 70% water in the sludge. = 4,969 BTU/kg
Total heating value from sludge is 0.00 MM BTU/hr
SLUDGE Value/year = $0

Sale of methane and lignin, based on BTU value is **$328,822 per year**

Total projected facility sales would be $29,676,765 per year

APITAL INVESTMENT ASSUMPTIONS

Total capital investment			
Civil Structural		(500,000)	
Area 100		6,146,434	
Area 200		14,955,166	
Area 300		4,028,307	
Area 307		3,714,334	
Area 400		651,440	
Area 500		7,515,486	
Area 600		9,824,251	
Area 700		234,910	
Area 800		3,684,612	
Area 900		2,236,491	
Fixed Capital		**$52,491,432**	
INDIRECTS Prorateable	3.5%	$1,837,200	
Process Development	2.0%	$1,049,829	
Field Expense	8.0%	$4,199,315	
Home Office Constr. Fee	12.0%	$6,298,972	
Contingency	10.0%	$5,249,143	
Start-up, Permits, Fees	3.0%	$1,574,743	
Working Capital per estimate		**$42,617,296**	1 mos Raw matls. + O&M
	Total Plant Cost	**$115,317,929**	
FEDERAL & STATE GRANTS	10%	($11,531,793)	
	Net Capital Investment	**$103,786,136**	

PERATING COST ASSUMPTIONS **8,400** hr/yr

Utilities (Rates based on **26,679,948** gal/yr produced)

	Amount/hr	*Units*	*$/unit*	*Cost /hr.*	**Total Cost /yr**
*Electricity	6,759	Kw-hr	$0.035	$237	$1,987,079
Well water	79,972	kg	$0.000	$0	$0
*Wastewater	39,119	kg	$0.00026	$10	$86,808
*Gypsum waste disposal	1,137	kg	$0.0364	$41	$347,327
		mTon	$1.103	$0	$0
Total Utilities				$288	$2,421,214

Quoted by High Plains

Model Input (purchased)

Raw Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
Corn Stover DRY (stm 101 less water)	37,500	kg	$0.680	$25,499.90	$214,199,143
*Sulfuric Acid (stm 710)	860	kg	$0.100	$86.26	$724,592
*Calcium Hydroxide (Lime stm 227)	337	kg	$0.293	$98.70	$829,039
*Ammonia (stm 717)	387	kg	$0.162	$62.77	$527,281
Corn Steep Liquor (stm 735)	708	kg	$0.051	$36.10	$303,280
Nutrients (stm 415)	0	kg	$0.291	$0.00	$0
Purchased Cellulase	14,021	lbs	$2.000	$28,041.97	$235,552,564
transport cost	750	miles	$3.000	$2250 /load	$48,684,298
*Natural Gasoline (stm 701)	391	kg	$0.155	$60.36	$506,988
*Rolling Stock Gasoline	79	kg	$0.155	$12.32	$103,470
*WWT Chemicals	5	kg	$0.000	$0.00	$0
*CW Chemicals	17	kg	$0.000	$0.00	$0
*BFW Chemicals	73.8	kg	$0.226	$16.65	$139,833
*Boiler Fuel (stm 813)	190	Mbtu	$2.500	$476.07	$3,998,989
Total Raw Materials				$54,391	$505,569,478

Quoted by High Plains

Processing Material Costs

	Amount/hr	Units	$/unit	Cost /hr.	Total Cost /yr
*Antifoam (Corn Oil)	0	kg	$0.304	$0	$0
Total Processing Materials				$0	$0

Quoted by High Plains

Operations and Maintenance Costs - DRY HANDLING (area 100)	each/day		wage	hr/day each	Total Cost /yr.
*Supervisors	0.5	$	20.00	12	$43,800
*Operators	2.0	$	16.00	12	$140,160
*Laborers	8.0	$	16.00	12	$560,640
*Maintenance	2.0	$	16.00	12	$140,160
Operations and Maintenance Costs - HYDROLYSIS/FERMENTATION (area 200, 300, 400, 500, 600)					
*Supervisors	1.0	$	20.00	12	$87,600
*Operators	8.0	$	16.00	8	$373,760
*Laborers	4.0	$	16.00	8	$186,880
*Technicians (Includes Lab.)	3.0	$	16.00	8	$140,160
*Maintenance	3.0	$	16.00	8	$140,160
Operations and Maintenance Costs - Utilities (area 700, 800, 900)					
*Supervisors	0.5	$	20.00	12	$21,900
*Operators	3.0	$	16.00	8	$70,080
*Laborers	1.0	$	16.00	8	$23,360
*Technicians	1.0	$	16.00	8	$23,360
*Maintenance	2.0	$	16.00	8	$46,720

Quoted by High Plains Standard HPY shifts are 12 hours.

Total Operations and maintenance labor costs $1,998,740

<u>Other Operations and Maintenance Costs</u>

Payroll Overhead	35% of operating labor	$	699,559
Maintenance Costs	2% of plant cost	$	1,049,829
Operating Supplies	0.25% of plant cost	$	131,229
Environmental	0.50% of plant cost	$	262,457
Local Taxes	1% of plant cost	$	524,914
Insurance	0.50% of plant cost	$	262,457
Overhead Costs	40% of labor, supervision,maint cost	$	799,496
Administrative Costs	1% of annual sales (less tax credits)	$	109,388
Distribution and Sales	0.5% of annual sales (less tax credits)	$	-
Total O&M Costs			$5,838,069

THER MODEL ASSUMPTIONS

verage prevailing market price of fuel grade ETOH:		$0.37	per kg
ssumes renewal of the ethanol excise tax credit of $.54 per gallon		$ 1.10	per gallon
nd the small producer tax credit of $.10 per gallon through the year 2007			
Value of CO_2 produced		$ 4.13	per metric ton
Price for Electricity		$ 0.035	per KWhr
Gas price per million BTU		$ 2.500	per MM BTU
		68%	Dry matter
orn Stover feedstock cost- **dry basis/short ton**	$ 14.45	$0.016	per kg
		$15.93	per metric ton
ant on-stream factor		0.959	
ant operating hours per year		8,400	
epreciable Life of Capital Equipment		15	years
verage annual commodity escalation rate:		3.0%	
verage annual cost escalation rate:		3.0%	

* ***Quoted by High Plains***

There are no land acquisiton costs included.
There are no off site costs included (e.g. public road improvements, extensions of power, water, telephone services)
There is a source of qualified construction personnel within daily driving distance of the site
There exist adequate roads and rail roads to allow equipment delivery.
The costs for air and water permits are not included.
Soils are adequate for conventional foundation designs.

CALCULATIONS FOR REQUIRED AMOUNT OF PURCHASED CELLULASE LIQUICELL 2500

ASED ON PUREVISION LABORATORY RESULTS OF COMPARISON

High Grade Waste Paper Substrate			
Soluble Charbohydrate % degraded in 18 hrs.			
Liquicell 2500	13%	87,059,020	ml/hr required for stover
PureVision Cellulase	82%	13,057,632	ml/hr required for stover
effectiveness multiple	6.43		

125	FPU/g protein	Liquicell 2500
731,295,772	liters/yr	Specialty
1.1000	S.G.	Enzymes
804,425,349	kg/yr	Inc.
193,062,084	gal/yr	
1,773,436,124	#/yr	
325,810	loads/yr	

0

cellulase storage tank

22,984	gal/hr
750,000	gal/vessel
33	vessel res. time (hr)

cellulase transfer pump

383	gpm

ASED ON PRODUCT SPECIFICATIONS PROVIDED BY SPECIALTY ENZYMES INC.

32	FPU/ml	Liquicell 2500
48,566,337	liters/yr	Specialty
1.1000	S.G.	Enzymes
53,422,971	kg/yr	Inc.
12,821,513	gal/yr	
117,776,282	#/yr	
21,637	loads/yr	

1

cellulase storage tank

14,021	gal/hr
750,000	gal/vessel
53	vessel res. time (hr)

cellulase transfer pump

234	gpm

ansport Calculations

10,000	lbs/axel		9.19	cellulase lb/gal
5	axels/truck		5,443	gal/truck
50,000	lbs/truck	$	**0.413**	**transport cost/lb**